
**Production of Poly EthyleneTerephthalate (PET)
resin from PTA and EG.
(BY POLYMERIZATION)**

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1. Abstract:

Polyester is a category of polymers which contain the ester functional group in their main chain. Although there are many polyesters, the term "polyester" as a specific material most commonly refers to polyethylene terephthalate (PET). Pure Terephthalic Acid (PTA) and Mono Ethylene Glycol (MEG) are mixed to produce oligomer and water in the Esterification section. The oligomer is then pumped to polymerization section after addition of chemicals namely catalyst, Antimony Tri Acetate. The oligomer is converted to polymer through polymerization reaction.

The proposed work in this project is the introduction to PET, its production process, global and local market analysis, plant study and material and energy balance of polyester plant.

2. Introduction

2.1. Polyethylene Terephthalate (PET)

PET is a polymer that possesses great importance in the contemporary world of plastics. Being a thermoplastic i.e. recyclable polymer made it the number one choice for numerous applications which satisfies the world need for a greener and more ecological alternative to commonly used plastics such as polyethylene and others.

Nowadays, Two PET grades dominate the global market fiber-grade PET and bottle-grade PET. They differ mainly in the end product properties such as optical appearance and production technologies where these properties can be controlled by molecular weight, intrinsic viscosity, and additives specific to each process or application. Other uses include film production and specialty nylons.

The scope of this report will focus on fiber-grade PET because of its high demand especially in the Pakistan market. The report discusses the historical development of PET, its importance, properties and material handling considerations.

2.2. Background

The raw materials to make PET are monoethylene glycol and purified terephthalic acid or dimethyl terephthalate. High purity is required of all raw materials.

In either case, the first step of the reaction is the formation of a pre-polymer, bis-hydroxyethyl terephthalate. Subsequent polymerization of this material with the removal of MEG forms the polymeric PET. The extent of polymerization conditions significantly affects the properties of resin being produced.

Until 1960s, DMT had been the preferred feed for PET manufacture, because the DMT was available in better pure form. With the development of high purity PTA, the free acid got acceptance and now used widely as feedstock. The use of PTA has eliminated the production and recovery needs of methanol and has added advantage that esterification to the prepolymer is considerably more rapid than the trans-esterification reaction, which is first step when starting from DMT.

Since due to adhesion addition quality, DMT process is still in commercial use in polyester film applications but due to economic benefits PTA has been mostly used feed in nowadays.

2.3. Innovation:

The improvement consists essentially in esterifying terephthalic acid and a glycol (ethylene glycol) in a mixture of lower molecular weight prepolymeric glycol terephthalates consisting substantially of ethylene glycol esters of terephthalic acid and wherein a molar ratio of esterified terephthalic acid to free and esterified (bound) ethylene glycol (total amount of glycol) in the reaction mixture is critical in that it must be at least 0.55. Esterification of the acid in the medium and in the proportions critical to the invention may be carried out at atmospheric pressure, and preferably at the boiling point of the mixture.

The invention has the advantages of

- (1) The formation of undesirable side-reaction products, such as ether esters, are avoided
- (2) The reaction rate is markedly increased
- (3) No methanol is evolved

3. Market and Competitive Analysis:

3.1. Purified Terephthalic Acid:

According to the data available in 2018, the global capacity of PTA production is 90 million tons per annum while the global annual production was 80.11 million tons in that year.

USA is the largest supplier of PTA in the world while Pakistan has only 1% contribution in global production of PTA. Asia-Pacific dominates the market demand for PTA by holding 77 percent of the total market share. The total market demand was 77.5 million tonnes. Among all the regions China holds the largest share of 57% in global market demand owing to its increasing industrialization, population, investments and growing demand from different sectors (construction, textile, beverages & packaging etc).

The major demand for PTA comes from Polyester fiber, films, and PET bottles. Among all the end-segment products Polyester fiber dominates the market demand by holding 65% of the market share in the global market demand.

In Pakistan, Lotte Chemical Pakistan Ltd is a world-class supplier of purified terephthalic acid, an essential raw material used in the polyester industry. The PTA plant was constructed in 1996/97 and started production in June 1998. Within a short time, PPTA's dedicated and highly motivated team of professional engineers proved that it could run this complex plant to world standards of safety, environmental care, product quality and process efficiency. Since 2002 the plant has operated above its nameplate capacity of 400,000 tons per annum and following minor de-bottlenecking and process improvements, is capable of ramping that up to 500,000 tons per annum.

3.2. Mono Ethylene Glycol:

The world MEG production capacity for the year 2016 was 34.8 million tons. The global production was dominated by Asian countries which accounted for 15.1 million tons from the total production.

Asia holds a prominent share in the global market production of MEG as it has a large number of fiber and textile producers which demand polyester and Polyethylene terephthalate. China, India & Taiwan are the leading producers of ethylene Glycol among other Asian-Pacific countries. Middle East Africa holds the second largest share in the production of MEG globally with the capacity of 11.9 million tons. The

production is high in Middle East Africa to meet their increasing demand of MEG in different industries including fiber, Anti-freeze, and packaging.

North America contributes 5.7 million tons in the global production which is followed by Europe with the contribution of 2.1 million tons.

Pakistan mostly imports MEG from Middle East and fulfill the requirements for the production of plastics.

3.3. PolyEthylene Terephthalete:

Global PET resin production capacity was 27.8 million tons in 2015. China is the leading producer with 27% of the market share followed by Europe (17%) & North America (17%). China is not only the largest producer of PET resin but also the largest consumer of PET bottles. In terms of region, Asia is the leading region in the production of Polyethylene Terephthalate (PET) resin with a market share of approximately 50% in 2015.

The other leading regions are Europe (17%) & North America (17%). In 2015 Asia region was a net exporter of PET resin. The top importer of PET resin among Asia region was Japan. In the last five year from 2010 – 2015, the export of PET resin grew by approximately 60% for the leading supplying countries of Asia region.

In Pakistan, Garton Industries and Ibrahim Fibers private limited are the major manufacturer of PET resin with the production capacity of 235,000 tons and 438,000 tons per annum respectively.

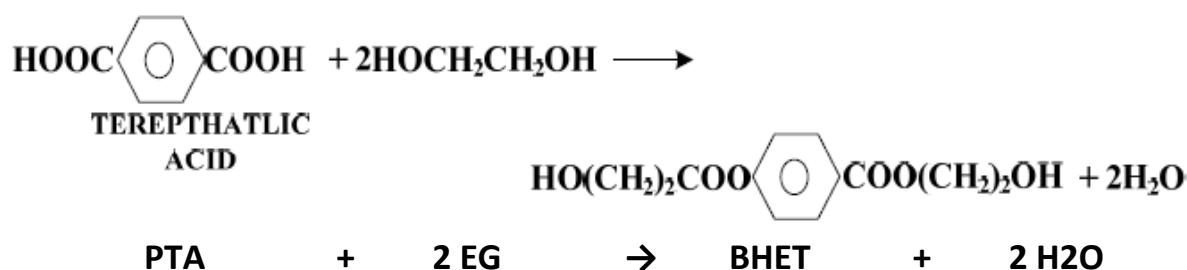
As markete demand is growing day by day due to increasing uses of plastics, so more plants of PET with greater production rate can be installed in Pakistan to increase the export and to meet the world's demand.

4. Process Description:

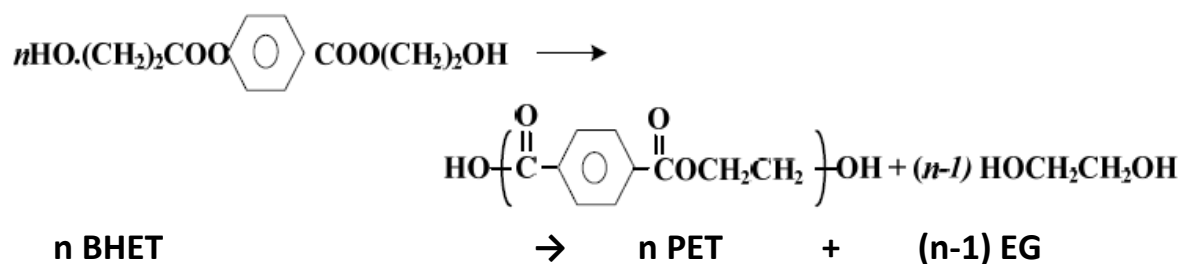
PET resins are produced commercially from ethylene glycol (EG) and either dimethyl terephthalate (DMT) or pure terephthalic acid (PTA). DMT and PTA are solids. DMT has a melting point of 140°C (284°F), while PTA sublimes (goes directly from the solid phase to the gaseous phase). But in this report we will focus on the production from pure terephthalic acid (PTA).

This process first produces the intermediate bi-(2-hydroxyethyl)-terephthalate (BHET) monomer and water. The BHET monomer is then polymerized under reduced pressure with heat and catalyst to produce PET resins. The reactions for the TPA process are:

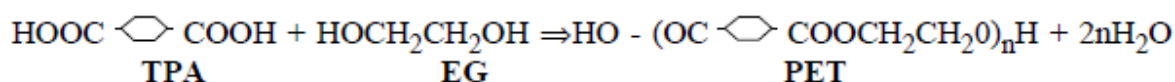
ESTERIFICATION REACTION



POLYCONDENSATION REACTION



OVERALL REACTION:



Overall process can be divided into five parts as follows

1. Paste Preparation.
2. Esterification Section
3. Pre-polycondensation
4. Polycondensation Section (DRR)
5. EG Recovery Section

4.1. Paste Preparation

In this unit, raw materials PTA and MEG is mixed properly in a paste tank under standard conditions. The tank is insulated so, the temperature is nearer to 55°C. Raw materials are mixed in a defined ratio given as, $ET=1.12$

By equation, 2 moles of MEG and 1 mole of PTA is required along with Antimony Triacetate catalyst to produce 1 PET. But in this process we are not giving 2 moles of MEG because the remaining MEG is taken from the MEG produced in the secondary reaction occurring with esterification. Catalyst is added to control mole ratio of EG and also to initiate polymerization.

PTA is continuously entering in the paste preparation vessel by a Schenk system. Schenk system measures PTA by load of PTA falling on the rotating wheel. MEG is fed with catalyst solution by spray nozzles in vent line of paste vessel. Control system controls the mole ratio of EG solution according to the quantity of PTA being added to the vessel. After suspending the paste by agitator the paste is pumped into Esterification reactor by positive displacement screw type pumps.

4.2. Esterification Section

At this stage, PTA reacts with MEG under high temperature and pressure to produce monomers of DGT (Diethylene-Glycol-Terephthalate) and split of water and polycondensation reaction is initiated.

Reaction occurs in CSTR. Vapors removed from the top of reactor contains SEG (spent ethylene glycol) and water. These vapors then enter in distillation column where EG is separated from water. EG removed is further used as EGR and water from distillation column moves toward condensers where it changes temperature and it is used as reflux.

4.3. Pre-Polycondensation

Product from ES is the fed to pre-polycondensation reactor. In CSTR monomers and oligomers undergo condensation polymerization and eliminate MEG. Antimony Triacetate (catalyst) is activated in this reactor and chain length initiate growth from here. The product monomers has two functional

groups (carboxyl and hydroxyl), it goes through step growth-polymerization, where the growth of molecules occur from the reaction of these two groups.

EG vapors from the top are captured by scrapper condenser which contains a cold EG shower to condense vapors. Scrappers are installed at the bottom of condenser which removes solid which may contaminate the walls of condenser. Liquid from the condenser is transferred to emulsion storage vessel which pump EG to coolers which is further used for EG showers. Level of vessel is set at 40% after which the EG is sent to distillation column. In column water is separated from EG.

4.4. DRR (Disc Ring Reactor)

DRR contains cage of rings attached to the shaft on the inner side of the reactor which continuously cuts the material. It is a horizontal reactor whose primary purpose is to achieve desired intrinsic viscosity up to 0.620. Material in DRR is thick and it may be called as semi-solid. The material is exposed to very high temperature and low pressure. Material slides through surface of every ring attached to shaft which is revolving at very slow rate per minute.

Similar to process in pre-polycondensation EG and water vapors leave from the top of DRR and pass towards horizontal scrapper condenser. Horizontal condenser is used because better quality product is required.

Uncondensed vapors pass through jet system afterwards it undergoes expansion which produces vacuum and condenses vapors. Some portion of uncondensed vapors vent to atmosphere which contains aldehyde and ketones,

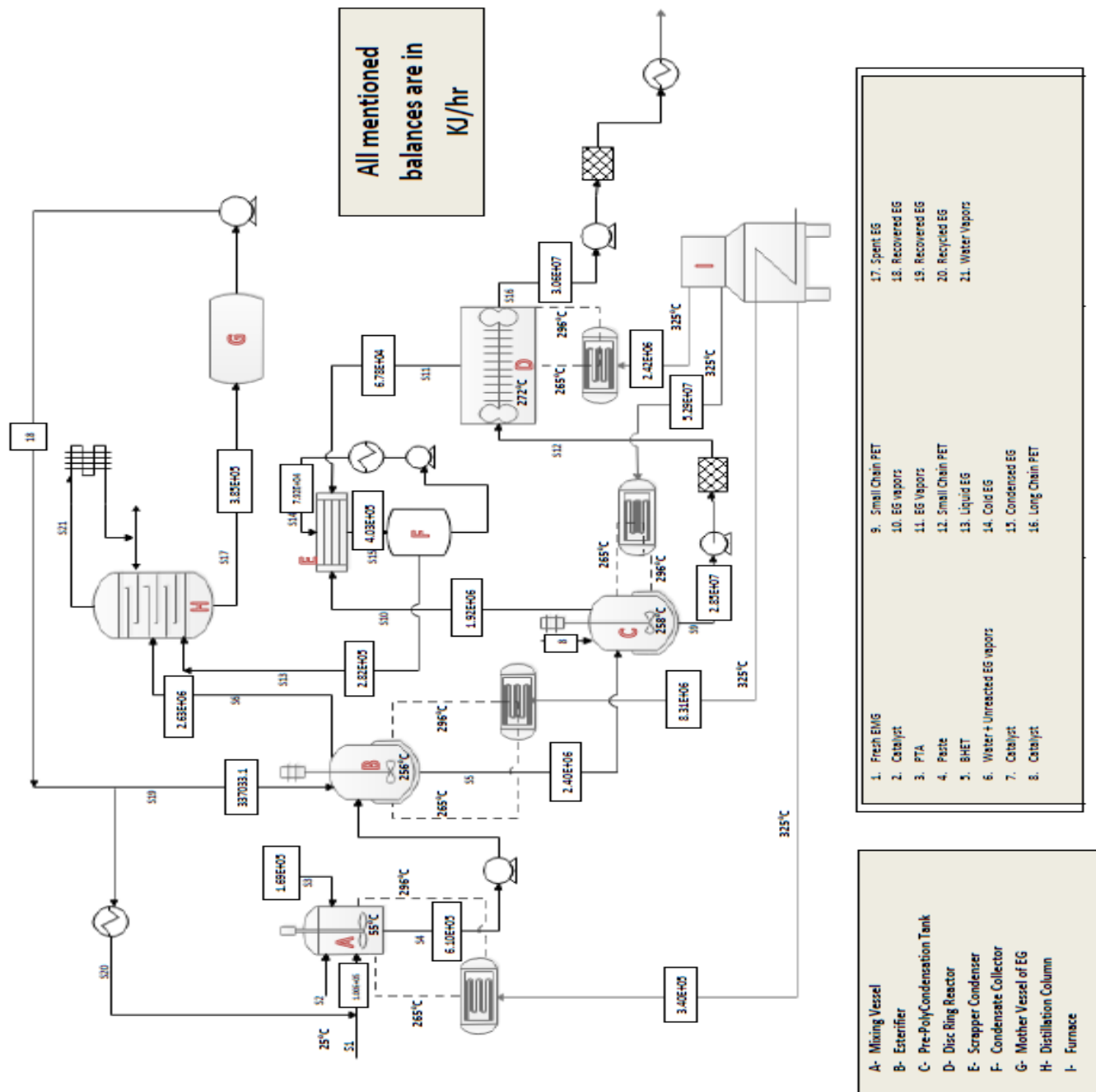
Product from DRR is pumped by pre-polymer pumps (product pumps which are gear type and self-lubricated) into viscometer and then to the candle filters which have stainless steel candles.

After filtration the product is then sent to drying section. From the outlet of DRR to the inlet in drying section it takes 20min and intrinsic viscosity at drying section will decrease up to 0.610 due to change in driving force.

4.5. EG Recovery Section

EG is recovered from distillation column and scrapper condenser. A distillation column is used for all reactors in the process to separate the EG leaving from reactors and scrapper condenser. In distillation column, ethylene glycol is recovered and used again in esterifier and water is condensed to use as reflux and rest is evaporated.

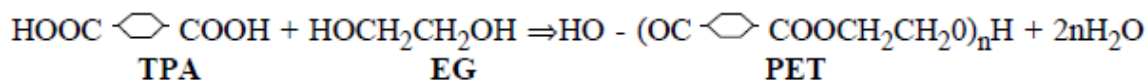
5. Process Flow Diagram:



6. Material and Energy Balance

6.1 Material Balance

Main Reaction



166.14 62.07 192.2 18

Basis = 100 Tonnes PET/day = 4166.6 kg PET/hr
 = 21.6781 mol of PET/hr

As overall conversion is 90% so

EG Required = 24.0876 mol of EG/hr = 1495.1 Kg EG/hr

PTA Required = 24.0876 mol of PTA/hr = 4001.9 Kg PTA/hr

6.1.1 Balance Across Mixer

Flowrate of Catalyst (Antimony TriAcetate) = 18 ppm

Since E/T = 1.12

Flowrate of MEG = 26.97801 Kmol/hr = 1674.59 kg MEG/hr

Fresh MEG = 24.08753 Kmol/hr = 1495.15 kg MEG/hr

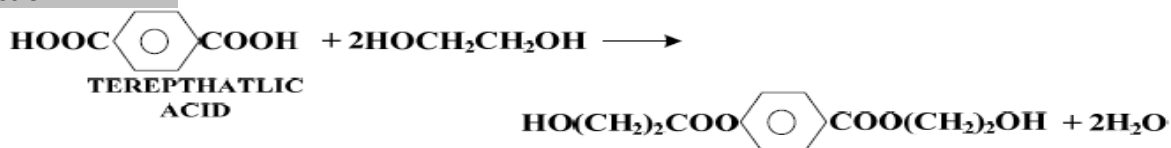
MEG from Recycle = 2.890506 Kmol/hr = 179.418 kg MEG/hr

Flowrate of PTA = 24.08753 Kmol/hr = 4001.98 Kg PTA/hr

	Input	Output
PTA	4001.908	
MEG	1674.529	
Catalyst		
Paste		5676.4366
Total	5676.437	5676.4366

6.1.2 Balance Across Esterifier

Reaction



166.14 62.07 254.24 18

	Input	Output	Stream
BHET in	5511.6198		5
un-reacted PTA	400.19077	400.1908	5
BHET unreactd		440.9296	9
PET poduced		3833.333	9
EG Vapors		1175.885	10
By Products		61.47166	10
Total	5911.81	5911.81	

6.1.4 Balance Across mother vessel of SEG

$$\begin{aligned} \text{Total EGR in mother vessel} &= \text{Stream 20} + \text{Stream 19} \\ &= 1495.115 \text{ kg MEG/hr} \end{aligned}$$

	Input	Output	Stream
SEG	1495.115		17
SEG		179.4138	19
SEG		1315.701	20
Total	1495.115	1495.115	

6.1.5 Balance Across Distillation Column

$$\begin{aligned} \text{input} &= \text{output} \\ S6+S13 &= S21+S17 \end{aligned}$$

$$X = 1196.092 \text{ kg/hr}$$

	Input	Output	Stream
EG(Un-reacted)	299.023		6
Water	780.437		6
EG Recovered	1196.092		13
Spent EG		1495.115	17
Water vapors		780.437	21
Total	2275.55	2275.55	

6.1.6 Balance Across DRR

$$\text{Unreacted BHET reacts} \quad 96\% = 423.2924 \text{ kg/hr} \quad 1.664932 \text{ mol/hr}$$

Unreacted BHET = 17.637183 kg/hr 0.069372 mol/hr

PET Formed = 320 kg/hr

Small chain PET input = 3833.3333 kg/hr

EG Vapors = 41.272352 kg/hr 0.664932 mol/hr

Long Chain PET = 4153.3333 kg/hr

input = output

PET in + BHET UR = Solids + EG Vapors + Long Chain PET

Solids= 62.02005 kg/hr

	Input	Output	streams
PET input	3833.333		12
BHET unreacted	440.9296	17.637183	12 & 16
Solids		62.020052	11
EG Vapors		41.272352	11
Long Chain PET		4153.3333	16
Total	4274.26	4274.263	

6.1.7 Balance Across Scraper Condenser

Total input of EG = S10 + S11 = 1217.158 kg/hr

Showering of Cold EG = X kg/hr

Condensed EG = Y kg/hr

un condensed Vapors out = Z kg/hr

Input = output
1083.028 + X = Y + Z (E1)

Distillate Collector

Input = output
Y = X + 1076.48 (E2)

Putting value of Y from E2 in E1 gives

$$\text{Non-condensed EG} = Z = 21.065573 \text{ kg/hr}$$

Since, 40% of tank is used in showering rest is sent to distillation column

So,

$$\text{EG Showered} = X = 797.39469 \text{ kg/hr}$$

From E1

$$\text{Condensed EG} = Y = 1993.4867 \text{ kg/hr}$$

Scrapper Condenser

	input	output	Streams
EG Vapors	41.27235		10
EG Vapors	1175.885		11
EG Showered	797.3947		14
Condensate		1993.487	15
Non-condensed		21.06557	
Total	2014.55	2014.55	

Condensate Collector

	input	output	streams
Condensate EG	1993.487		15
Cold EG		797.3947	14
Liquid EG		1196.092	13
Total	1993.49	1993.49	

6.2 Energy Balance

Constants For Cp (KJ/Kmol.K)

	A	B	C	D		E
BHET	412.74	-0.12496	0.00020321	0		
w (vap.)	32.243	0.0019238	1.0555E-05	-3.6E-09		
PTA	364.2	-0.61239	0.0011058	0		
EG (vap.)	35.697	0.24832	-0.0001497	3.01E-08		
EG (liq.)	35.54	0.43678	-0.0001849	0		
PET	9156.4	-3.5236	-0.0068921	0		
w (liq.)	276.37	-2.0901	0.008125	-0.000014116		9.37E-09
SEG (liq.)	513.42	-0.6519	-0.00006	0.0000009		
SEG (vap.)	73.06	0.34441	0.0001468	1.846E-08		
Diphenyl ether	956.86	-5.559	0.009612			

Where

Cp =

6.2.1 Balance Across Mixer

Conditions:

T=328K

P=1 atm

Input	(At standard Temp)				
Substance	Flowrate (Kmol/hr)	Cp (KJ/Kmol.K)	ΔT (K)	Q= $mCp\Delta T + m\lambda$ (KJ/hr)	Streams
PTA	24.0875631	2.80E+02	25	1.69E+05	3
MEG	26.9780707	1.49E+02	25	1.00E+05	1+20
Total				2.69E+05	

OutPut	(At 55°C = 328K)				
Substance	Flowrate (Kmol/hr)	Cp (KJ/Kmol.K)	ΔT (K)	Q= $mCp\Delta T + m\lambda$ (KJ/hr)	Streams
PTA	24.0875631	2.82E+02	55	3.74E+05	4
MEG	26.9780707	1.59E+02	55	2.36E+05	5
Total				6.10E+05	

As

Qin	-	Qout	=	Net Heat
Net Heat		=	-3.40E+05	KJ/hr

Flowrate of Dow Therm Required for this Heat

$$T = 286^{\circ}\text{C} \quad 559 \text{ K}$$

$$C_p = 842.61 \text{ KJ/Kmol.K}$$

As

$$Q = mC_p\Delta T$$

$$M = 20.2000378 \text{ Kmol/hr}$$

6.2.2 Balance Across Esterifier

Conditions:

$$T=256^{\circ}\text{C} \quad \text{or } 529\text{K} \quad P=0.14\text{bar}$$

Input	(At 328K)				
Substance	Flowrate	Cp	ΔT	$Q=mC_pT$	Streams
	(Kmol/hr)	(KJ/Kmol.K)	(K)	(KJ/hr)	
PTA	24.08756311	2.82E+02	55	3.74E+05	4
MEG	26.97807068	1.59E+02	55	2.36E+05	4
Makup-EG	21.19705554	159	100	337033.18	19
Total				9.47E+05	

OutPut	At 529 K					
Substance	λ	Flowrate (Kmol/hr)	Cp (KJ/Kmol.K)	ΔT (K)	Q= mCp ΔT +m λ (KJ/hr)	Streams
	KJ/Kmol					
BHET		21.6788	4.04E+02	256	2.24E+06	5
EG vap (un-reacted)	6.56E+04	4.8175	1.30E+02	256	4.76E+05	6
PTA(un-reacted)		2.4088	3.50E+02	256	2.16E+05	5
Water (vap)	4.07E+04	43.3576	3.57E+01	256	2.16E+06	6
				Total	5.09E+06	

Heat of reaction

Reactants

PTA	-815963.68	KJ/Kmol
MEG	-384852.6	KJ/Kmol

Products

BHET	-1094324	KJ/Kmol
Water	-241562	KJ/Kmol

$$H_r = -4160033 \text{ KJ/hr}$$

$$Q_{in} - Q_{out} + H_r = \text{Net Heat}$$

Net Heat	=	-8.31E+06 KJ/hr
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Flowrate of Dow Therm Required for this Heat

(Dow therm is Diphenyl Ether in vapor form)

$$T = 284^{\circ}\text{C} \quad 557 \text{ K}$$

$$C_p = 842.610388 \text{ KJ/Kmol.K}$$

As

$$Q = mC_p\Delta T$$

m	=	Q/CpΔT
m	=	352.073779 Kmol/hr

6.2.3 Balance Across Pre-Polycondensation Tank

Conditions:

T=272°C or 545K P=19 – 20mbar

Input	At	529 K				
Substance		Flowrate (Kmol/hr)	Cp (KJ/Kmol.K)	ΔT (K)	Q= mCpΔT+mλ (KJ/hr)	Streams
BHET in		21.6788068	4.04E+02	256	2.24E+06	5
un-reacted PTA		2.4088	3.50E+02	256	2.16E+05	5
Total					2.46E+06	

OutPut	At	545 K				
Substance	λ (KJ/Kmol)	Flowrate (Kmol/hr)	Cp (KJ/Kmol.K)	ΔT (K)	Q= mCpΔT+mλ (KJ/hr)	Streams
un-reacted PTA		2.4088	3.59E+02	272	2.35E+05	9
BHET unreactd		1.73430454	4.05E+02	272	1.91E+05	9
PET poduced		19.9445023	5.19E+03	272	2.82E+07	9
EG Vapors	6.56E+04	18.9445023	1.31E+02	272	1.92E+06	10
Total					3.05E+07	

Heat of reaction

Reactants

BHET	-1094324	KJ/Kmol
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Products

PET	-1974328	KJ/Kmol
MEG	-384852.6	KJ/Kmol

Hr	=	-24836388 KJ/hr
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Qin - Qout + Hr = Net Heat

Net Heat	=	-5.29E+07 KJ/hr
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Flowrate of Dow Therm Required for this Heat

(Dow therm is Diphenyl Ether in vapor form)

DowTherm Conditions
T = 284°C 557 K

$$C_p = 842.610388 \text{ KJ/Kmol.K}$$

As

$$Q = m C_p \Delta T$$

$$m = Q / C_p \Delta T$$

m	=	5229.55818 Kmol/hr
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6.2.4 Balance Across DRR

Conditions:

T=548K

**P= 1.5-1.2
mbar**

Input	AT	545 K					
Substance	Flowrate (Kmol/hr)	λ KJ/Kmol	C_p (KJ/Kmol.K)	ΔT (K)	$Q = m C_p \Delta T + m \lambda$ (KJ/hr)	Streams	
BHET unreacted	1.7343045		4.05E+02	272	1.91E+05	12	
PET	19.944502		5.19E+03	272	2.82E+07	12	
Total					2.83E+07		

OutPut	(At 272°C)	548 K					
Substance	Flowrate (Kmol/hr)	λ KJ/Kmol	C_p (KJ/Kmol.K)	ΔT (K)	$Q = m C_p \Delta T + m \lambda$ (KJ/hr)	Streams	
Unreacted BHET	0.0693722		4.05E+02	275	7.73E+03	16	
Long Chain PET	21.609435		5.16E+03	275	3.07E+07	16	
EG vapors	0.6649324	6.56E+04	1.32E+02	275	6.78E+04	11	
Total					3.07E+07		

$$Q_{in} - Q_{out} = \text{Net Heat}$$

$$\text{Net Heat} = 2.39E+06 \text{ KJ/hr}$$

Flowrate of Dow Therm Required for this Heat

(Dow therm is Diphenyl Ether in vapor form)

$$T = 286^\circ\text{C} \quad 559 \text{ K}$$

$$C_p = 842.61 \text{ KJ/Kmol.K}$$

As

$$Q = m C_p \Delta T$$

m	=	258.17667 Kmol/hr
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6.2.5 Balance Across Distillation Column

Conditions:

P= 260
mbar

Input						
Substance	Flowrate	λ	Cp	ΔT	Q= $mCp\Delta T+m\lambda$	Streams
	(Kmol/hr)	KJ/Kmol	(KJ/Kmol.K)	(K)	(KJ/hr)	
EG vap(Un-reacted)	4.8175126	6.56E+04	1.30E+02	256	4.76E+05	6
Water vapors	43.357614	4.07E+04	3.57E+01	256	2.16E+06	6
EG liq Recovered	19.27005		1.83E+02	80	2.82E+05	13
Total					2.92E+06	

OutPut						
Substance	Flowrate	λ	Cp	ΔT	Q= $mCp\Delta T+m\lambda$	Streams
	(Kmol/hr)	KJ/Kmol	(KJ/Kmol.K)	(K)	(KJ/hr)	
EG Liq	24.087563	6.56E+04	1.60E+02	100	3.85E+05	17
Condensed Water	43.357614	4.07E+04	6.50E+01	40	1.13E+05	20
Total					4.98E+05	

Net Heat = 2.42E+06 KJ/hr

This is energy generated which can be used in utility section

6.2.6 Balance Across Scrapper Condenser

Input						
Substance	Flowrate	λ	Cp	ΔT	Q= $mCp\Delta T+m\lambda$	Streams
	(Kmol/hr)	KJ/Kmol	(KJ/Kmol.K)	(K)	(KJ/hr)	
EG Vapors	0.6651467	6.56E+04	1.32E+02	275	6.78E+04	11
EG Vapors	18.950608	6.56E+04	1.31E+02	275	1.93E+06	10
EG Showered	12.850841		1.54E+02	40	7.92E+04	14
Total					2.07E+06	

Output						
Substance	Flowrate	λ	Cp	ΔT	Q= $mCp\Delta T+m\lambda$	Streams
	(Kmol/hr)	KJ/Kmol	(KJ/Kmol.K)	(K)	(KJ/hr)	
Condensed EG	32.116751		1.57E+02	80	4.03E+05	15

Qin - Qout = Net Heat

Qnet = 1.67E+06

7. Equipment Design

7.1 DESIGN OF MIXING VESSEL

7.1.1. INTRODUCTION

Mixing is an essential operation in many engineering fields. It has central significance in food processing, pharmaceutical production, chemical engineering, biotechnology, agri-chemical preparations, paint manufacturing, water purification among countless other applications. One of the main aims of agitation systems using stirred vessels is to maintain balanced quantities of substances in different phases based on concentration levels. In cases where soluble solids are mixed, stirrers are used to increase interaction between the particles and avoid uneven accumulation at one point.

In large-scale mixing plants, stirrers and the entire agitation set-up should be able to create faster movement of substances and high turbulence. This makes the entire mixing process in large containers complicated and impractical to study through experiments. There is therefore a need to provide a more workable method that will simplify the process.

7.1.2. BASIC MIXING EQUIPMENT DESIGN

The dimensions of liquid content of the vessel, and dimensions and arrangement of impellers, baffles and other internals are factors that influence the amount of energy required for achieving a needed amount of agitation or quantity of mixing. The internal arrangements depends on the objectives of the operation: whether it is to maintain homogeneity of a reacting mixer or to keep a solid suspended or a gas dispersed or to enhance heat or mass transfer. A basic range of design factors, however, can be defined to cover the majority of cases.

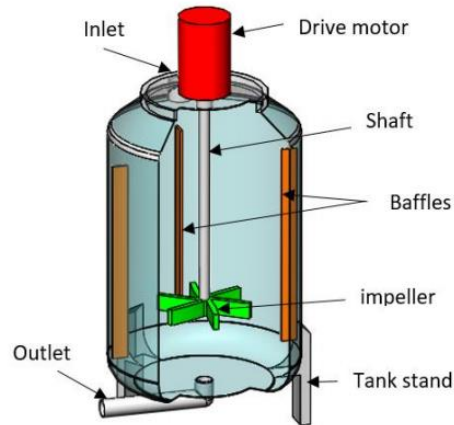


Figure 1. Experimental apparatus set-up showing the assembly that was used to perform the mixing experiment.

1. PARAMETERS

- Viscosity
- Density
- Temperature & Pressure
- Blend time
- Volume
- Any specific process requirement.

2. IMPELLERS

The function of any mixing impeller is to convert the rotational energy of the mixer shaft into the correct combination of flow, shear and turbulence to achieve the required process results.

As no one-impeller design is capable of providing optimum performance under every process condition, optimum process performance is dependent upon selecting an impeller design that has the specific characteristics required by a given process.

• NUMBERS OF IMPELLERS

The use of a single impeller is usual preferred option on the basis of cost. However, changes in the ratio of liquid level (Z) to vessel diameter (T) can have an adverse effect on the flow pattern generated within the vessel. This can result in the need to consider the use of multiple impeller in order to achieve an economic solution.

Z/T ratio alone is not only consideration when determining the number of impellers required. Multiple impellers may also be considered for other reasons including, when high viscosity fluids are involved, for mixing at low level during filling and emptying or where draw down from the liquid surface is a requirement.

• IMPELLER POSITIONING

Whether utilizing a single or multiple impeller configuration, the positioning of the impellers within the process fluid can have a significant effect on the overall process

performance. Incorrect poisoning can lead to staged flow pattern, poor dispersion of additives and impeller being out of the liquid at crucial stages of the process.

3. D/T RATIO

The ratio of mixing impeller diameter (D) to vessel diameter (T) has a very significant effect on the performance of the most fluid mixers and the optimum D/T is a function of both process conditions and process requirement.

Normally the optimum D/T will be in the range of $0.2 < D/T < 0.5$. Some special applications however, sometimes operate outside this range.

4. BOTTOM CLEARANCE

The impeller bottom clearance (C/T ratio) can also have a very significant effect on the overall performance of a mixer, effecting both power dear and pumping efficiency. The optimum C/T ratio essentially dependent upon impeller type but can also be effected by process conditions.

Normally the optimum C/T will be in the range of $0.1 < C/T < 0.3$. Hydrofoils however can operate at much higher levels, up to $C/T = 0.5$ or more.

5. VESSEL DESIGN

When designing a vessel for mixer duty it is important to understand the role that tank geometry plays in determining the final mixer design. Poor aspect ratios and inappropriate bottom shapes can both result in increases mixer cost and in certain circumstances make it impossible to optimize the mixer design.

6. ASPECT RATIO

It is generally accepted that the ideal aspect ratio for most mixing tank is one where the liquid depth (Z) is equal to the tank diameter (T) as this allows for the optimum number of impeller, optimum power input and optimum power distribution.

In practice, the optimum Z/T will be in the range of $0.9 < Z/T < 1.2$ as this does not significantly effects mixer design.

7. BOTTOM SHAPE

Tank bottom shape can have a significant effect on the flow patterns generated within the mixing vessel and hence the mixer ability to achieve optimum process performance. Normally a dish-bottom tank is the preferred bottom shape. However, the flat-bottom and shallow cones (less than 15°) can be used for many processes without any particular problem. In the case of flat-bottomed tanks mixer performance can often be significantly improved by the introduction of corner fillets. In general deep cones should be avoided especially where the requirement is solid suspension.

8. BAFFLES

The importance of proper tank baffling in obtaining mixer performance should not be underestimated. In a correctly baffled tank the mixer develops the fluid regime required to achieve the optimum process results.

An incorrectly baffled tank on the other hand can lead to poor mixer performance and may even result in the mixer not being able to achieve the process result for which it was designed.

It is normally desirable to set the baffles off the wall and off the bottom of the tank to prevent solids or fluids from stagnating at those points. The optimum baffling arrangement however, will vary from process to process and is dependent upon a variety of factors including, vessel internals, specific power, the required surface effect and viscosity.

DESIGN STEPS AND CALCULATIONS

- Volume of vessel
- Height and diameter of vessel
- Turbine design
- Shape factor
- Rotational speed
- Residence time
- Reynold's number
- Power calculation

9.DESIGN PROCEDURE

The proposed designing specifications of the tank meant for preparing the paste of ethylene glycol and pure terephthalic acid of the desired consistency, the mole ratio of 1:2 are evaluated below.

1. FEED DATA

- Mixing temperature: 55°
- Mixing time: 2hr

COMPONENT	MOLE FRACTION	DENSITY	VISCOSITY
EG	0.88	1092.218	0.0069724
TPA	0.12	1509.9995	6.6733*10 ⁻⁶

2. PASTE PROPERTIES AT 55 °C

- PASTE DENSITY

$$\begin{aligned}
 P_T &= (x * \rho) + (x * \rho) \\
 &= (0.88*1092.218) + (0.12*1509.9995) \\
 &= 961.152 + 181.2 \\
 &= 1142.35194 \text{ kg/m}^3
 \end{aligned}$$

- PASTE VISCOSITY

$$\mu_T = (x * \mu) + (x * \mu)$$

$$\begin{aligned}
 &= (0.88 \times 0.0069724) + (0.12 \times 6.6733 \times 10^{-6}) \\
 &= 6.135712 \times 10^{-3} + 8.00796 \times 10^{-7} \\
 &= 6.1365 \times 10^{-3} \text{ Pa}\cdot\text{sec}
 \end{aligned}$$

3. FLUID CHARACTERISTICS IN MIXING VESSEL

- FLOW RATE

$$\begin{aligned}
 \text{Total mass flow rate (out)} &= m = 5676.4366 \\
 \text{Volumetric flow rate} &= Q = m/\rho_T = 5676.4366/1142.35194 \\
 &= 4.969 \text{ m}^3/\text{hr}
 \end{aligned}$$

- PASTE VOLUME

$$\begin{aligned}
 V &= Q \times t = 4.969 \times 2 \\
 &= 9.938 \text{ m}^3
 \end{aligned}$$

➤ **Vessel Dimension**

Since, 60% of the total vessel volume is filled with the fluid so the total volume of the vessel will be estimated as:

$$\begin{aligned}
 \text{Total Mixer Capacity} &= 9.938/0.6 \\
 &= 16.56 \text{ m}^3
 \end{aligned}$$

- **Vessel Height , Diameter & Volume**

RELATIONS	HEIGHT (m)	DIAMETER (m)	VOLUME (m ³)
H=3D	5.56	1.853	15.00
H=2D	4.24	2.12	14.97
H=D	2.67	2.67	14.95

Selected Dimensions

- Height H = 5.56
- Diameter D = 1.853

2. Liquid Levels in the vessel:

As 60% of the total volume is paste filled having volume = 9.938

So, the height of the liquid in vessel is:

$$\begin{aligned}
 \Rightarrow \text{Height /Level of the Liquid} &= h = \text{Volume} / \pi r^2 \\
 &= 9.938 / (3.142 \times 1.853^2 / 4) \\
 &= 3.687 \text{ m}
 \end{aligned}$$

Since, 65% of paste is above agitator & 35% is below it.

$$\text{Level above agitator} = 0.65 * 3.687$$

$$= 2.39658 \text{ m}$$

$$\text{Level below agitator} = 0.35 * 3.687$$

$$= 1.29045 \text{ m}$$

4. Impeller selection

6 pitched-blade turbine type axial flow impeller is selected as an agitator (impeller in which the blade makes an angle of less than 90° with the plane of rotation), as a top-entering agitator shafts. Four baffles are estimated as optimum along with the turbine type impeller for effective mixing (also because the Reynolds number of the paste is greater than 2000).

Proportion of tank

$$\text{At } H = 5.56 \text{ m}$$

$$D = 1.853 \text{ m}$$

✓ Impeller diameter

$$\frac{\text{impeller diameter}}{\text{tank diameter}} = \frac{D_a}{D} = \frac{1}{3}$$

$$D_a = (1/3)(1.853)$$

$$D_a = 0.618 \text{ m}$$

✓ Baffle width:

$$\frac{\text{baffle width}}{\text{tank diameter}} = \frac{J}{D} = \frac{1}{10} \quad (\text{European standard})$$

$$J = (1/10)(1.853)$$

$$J = 0.1853 \text{ m}$$

✓ Clearance between baffle and vessel wall

$$(0.1-0.15) J$$

$$C_a = 0.1J$$

$$C_a = 0.1 \times 0.1853 \text{ m}$$

$$C_a = 0.01853 \text{ m}$$

✓ Impeller height from bottom

$$\frac{\text{impeller height from bottom}}{\text{impeller diameter}} = \frac{C}{D_a} = 1$$

$$C = 0.618 \text{ m}$$

✓ Impeller blade width

$$\frac{\text{width of impeller blade}}{\text{impeller diameter}} = \frac{W}{Da} = \frac{1}{15}$$

$$W = (1.853/15)$$

$$W = 0.1235 \text{ m}$$

✓ Blade length

$$\frac{\text{length of impeller blade}}{\text{impeller diameter}} = \frac{L}{Da} = \frac{1}{4}$$

$$L = (1/4) (0.618)$$

$$L = 0.1545 \text{ m}$$

Fluid flow characteristics:

✓ Velocity of fluid in tank

$$V = \frac{Q}{A}$$

$$A = \pi r^2 = (3.14)(1.853/2)^2$$

$$A = 2.6954 \text{ m}^2$$

$$V = (4.969/2.6954)$$

$$V = 5.12086 \text{ (m/sec)}$$

✓ Reynolds number

$$Re = \frac{(\rho u d)}{\mu}$$

$$Re = \frac{(1142.35194)(5.12086)(1.853)}{0.0061365}$$

$$Re = 1766434$$

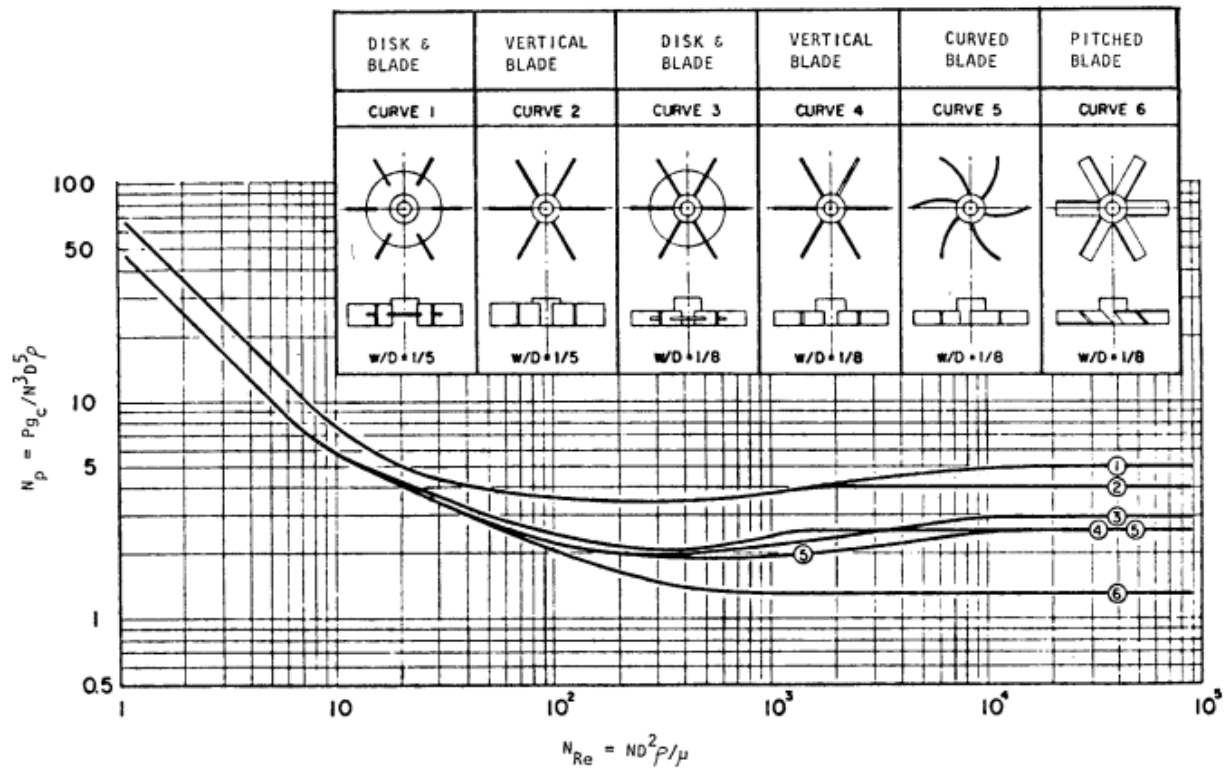
✓ Impeller rotational speed

$$N = \frac{(Re)(\mu)}{(Da)(Da)(\rho)}$$

$$N = \frac{(1766434)(0.0061365)}{(0.618)(0.618)(1142.35194)}$$

$$N = 24.845 \text{ rps}$$

$$N = 1490 \text{ rpm}$$



✓ Power consumption by impeller

Estimating the power consumption for any agitator is essential to design. The power consumed by the agitator depends on fluid density, viscosity, vessel dimension, external attachment, position of impeller, its rotational speed, its shape and impeller diameter by plot of power number versus Reynolds number. Thus, the empirical relation for calculating power is given by:

$$P = N_p \rho n^3 D_a^5$$

$$P = (1.3)(1142.35194)(25)^3(0.618)^5$$

$$P = 2091730 \text{ watt}$$

$$P = 2091.730 \text{ kW}$$

Specification sheet:

ENTITY	PROJECTED DIMENSION
mixing temperature	55 °C
mixing time	2 Hrs
viscosity of paste	6.1365E-3 Pa.sec
density of paste	1142.35194 kg/m ³
mixer capacity	16.56 m ³
vessel height	5.56 m
vessel diameter	1.853 m
liquid level	3.687 m
level above agitator	2.39658 m
selected impeller	6 blade pitched turbine type
impeller diameter	0.618 m
number of baffles	4
baffle width	0.1853 m
baffle clearance	0.01853 m
impeller bottom clearance	0.618 m
impeller blade width	0.1235 m
blade length	0.1545 m
impeller rotational speed	1490 rpm
power consumption by impeller	2091.730 kW

7.2. Design of Esterification Tank (Reactor)

Types and example of reactor:

In chemical manufacturing operations, the reactor is the central equipment item of the plant. There are two basic vessel types of chemical reactors, named as tank and pipe. They are both used in a batch or continuous mode. The reactor is the heart of a chemical process. It is the only place in the process where raw materials are converted into products.

Chemical process equipment selection and design, 3rd edition by J.R.Couper, W.R.Penny, J.P.Fair, S.M.Walas, page no 608

In reactor, chemical reactions are carried out to convert raw material to product. Reactors are usually designed as special items for a project. Reactors are unique in nature when its system involve catalysis and multiphases. Even conventional stirred tanks are often customized for the application by optimization of internal mixing pattern, heat transfer surface, or control instrumentation, feed addition points.

Reactor design is an important step in the overall design of the process, and it is important that the reactor will be capable of achieving the desired yields and selectivity when operated at full scale. Few industrial reactors are designed based on detailed kinetics and hydrodynamics data ; instead, reactors are usually scaled up from pilot plant reactors, making suitable allowance for heat transfer, residence time, mass transfer, or whatever other parameters have been determined to manage the size. The final volume of real reactors is often determined more by the needs for mixing, separation, or heat transfer than by the residence time required for the reaction.

Chemical Engineering Design, Principle , practice and Economics of plant and process design, second edition by Gavin Towler, Ray Sinnott , page no 631

Design of industrial chemical reactor:

The design of an industrial chemical reactor must satisfy the following requirements:

- The chemical factors: the kinetics of the reaction. The design must provide adequate residence time for the desired reaction to proceed to the required degree of conversion.
- The mass transfer factors: with heterogeneous reactions the reaction rate may be controlled by the rates of diffusion of the reacting species; than that of chemical kinetics.
- The heat transfer factors: the removal, or addition, of the heat of reaction.
- The safety factors: the confinement of dangerous reactants and products, and the control of the reaction and the process conditions.

Principal types of reactor:

Mode of operation	Phases present	Reactor geometry
• Batch • continuous	• homogeneous • heterogeneous	• stirred tank reactor • tubular reactor • packed bed, fixed and moving; • fluidised bed.

1. Stirred tank reactors:

Stirred tank (agitated) reactors consist of a tank fixed with a mechanical agitator and a cooling jacket or coils. They are operated as batch reactors or continuously. Several reactors may be used in series.

The stirred tank reactor can be considered the basic chemical reactor; modelling on a large scale the conventional laboratory flask. Tank sizes range from a few litres to several thousand litres. They are used for homogeneous and heterogeneous liquid-liquid and liquid-gas reactions; and for reactions that involve finely suspended solids, which are held in suspension by the agitation. As the degree of agitation is under the designer's control, stirred tank reactors are mainly suitable for reactions where good mass transfer or heat transfer is necessary.

When operated as a continuous process the composition in the reactor is constant and the same as the product stream, and, except for very rapid reactions, this will limit the conversion that can be obtained in one stage. The power requirements for agitation will depend on the degree of agitation required and will range from about 0.2 kW/m³ for moderate mixing to 2 kW/m³ for intense mixing.

2. Tubular reactor:

Tubular reactors are generally used for gaseous reactions, but are also suitable for some liquid-phase reactions. If high heat-transfer rates are required, small-diameter tubes are used to increase the surface area to volume ratio. In tubular reactor several tubes are arranged in parallel way same as that of shell and tube heat exchanger. For high-temperature reactions the tubes may be arranged in a furnace.

3. Packed bed reactors:

There are two basic types of packed-bed reactor: one in which the solid is a reactant, and one in which the solid is used as a catalyst. Examples in which solid is used as a reactant include extractive metallurgical industries.

In the chemical process industries the designer will normally be concerned with the second type: catalytic reactors. Industrial packed-bed catalytic reactors range in size from small tubes, a few centimetres diameter, to large diameter packed beds. Packed-bed reactors are used for gas and gas-liquid reactions. Heat-transfer rates in large diameter packed beds are poor and where high heat-transfer rates are required fluidised beds should be considered.

4. Fluidised bed reactors:

The essential features of a fluidised bed reactor is that the solids are held in suspension by the upward flow of the reacting fluid; this promotes high mass and heat-transfer rates and good mixing. Heat-transfer coefficients in the order of $200 \text{ W/m}^2\text{°C}$ to jackets and internal coils are typically obtained. The solids may be a catalyst; a reactant in fluidised combustion processes; or an inert powder, added to help heat transfer.

Though the principal advantage of a fluidised bed over a fixed bed is the higher heat transfer rate, fluidised beds are also useful where it is necessary to transport large quantities of solids as part of the reaction processes, such as where catalysts are transferred to another vessel for regeneration.

Fluidisation can only be used with relatively small sized particles, $< 300 \text{ }\mu\text{m}$ with gases. A great deal of research and development work has been done on fluidised bed reactors.

Design steps and procedure:

- Volume is calculated with the help of
 - Rate of reaction
 - Residence time
- Then power consumed is calculated.
- Mechanical design
 - Thickness of shell
 - Impeller design
 - Diameter of impeller

Procedure:

- Collect all the kinetic and thermodynamic data on the desired reaction.
- Collect physical data for design.
- Identify rate controlling mechanism.
- Choose a suitable type of reactor.
- Make an initial selection of reactor condition to give desired conversion and yield.
- Size the reactor.
- Make a preliminary design of reactor.

General selection Criteria of reactors:

The selection of the best reactor type for a given process is subject to the number of major considerations. Such design aspects include:

- Temperature and pressure of reaction.
- Need for removal of reactants and products.
- Required pattern of product delivery.
- Catalyst and considerations, such as requirement for solid catalyst particle replacement and contact with fluid reactants and products.
- Cost of the reactor.

Plant design and economics for chemical engineers, by M.S.Peter, K.D.Timmerhaus, R.E.West , 5th edition page no 621

Esterification reactor:

A continuous stirred tank reactor has been designed to work as the esterification reactor.

Several types of reactors are commonly used for carrying out esterification reactions, each have their own advantages and characteristics. The following tables summarized their advantages and disadvantages.

Plug Flow Reactor (PFR)		continuous stirred tank reactor (CSTR)	
advantages	disadvantages	advantages	disadvantages
high Reynold's number (turbulent flow)	lack of agitation	overcomes mixing problems	wide residence time distribution
uniform velocity across the reactor	problems with high viscous products	good heat transfer characteristics	cannot attain high conversion
		controlled deposit using appropriate stirrer	wall using

CSTR is selected because of following reasons:

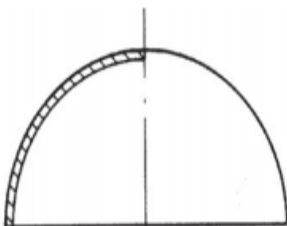
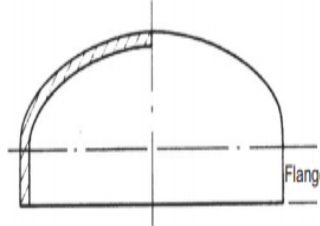
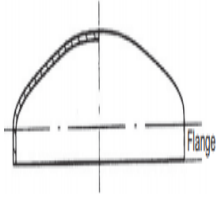
- When CSTR is used in series the degree of polymerization increases.

- Increasing the reactor size and the number of reactors allow high conversions for given molecular weight which would reduce recovery cost for monomers.
- For large volume.
- Low operating cost.
- High output rate.
- More uniform product quality.
- Simplicity of operation.
- High mass and heat transfer efficiency.

Head and closure of reactor:

Head and closure used in vessel are of following types:

- Hemispherical
- Ellipsoidal
- Torispherical

characteristics	Hemispherical	Ellipsoidal	torispherical
			
pressure bearing	strong in pressure bearing	Moderate pressure can be beard	moderate pressure can be beard
Assembly	Very difficult to join with shell with the help of bearing. extra sealing are required to prevent leakage.	easy joining with shell with the help of flanges. extra sealing is not required.	easy joining with shell with the help of flanges. extra sealing is not required.
Agitator assembly	very difficult to avoid agitator from fluctuation	very easy to maintain agitator on a constant speed without fluctuation	a bit difficult maintenance than ellipsoidal head
cost of formation	most expensive	less expensive	less expensive
thickness calculation	$t = \frac{PiDi}{4SE - 0.4Pi}$	$t = \frac{PiDi}{2SE - 0.2Pi}$	$t = \frac{0.885PiRc}{SE - 0.1Pi}$

s

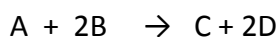
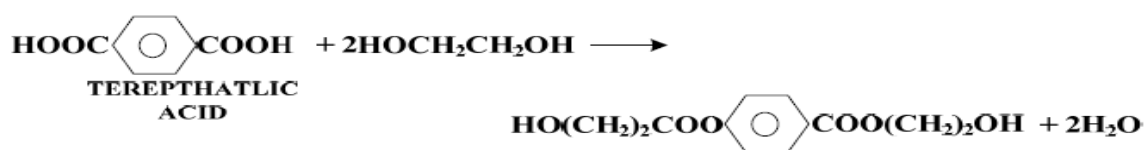
Design calculations:

The design equation of CSTR is

$$V/F_{AO} = X_A / (-r_A)$$

Chemical Reaction Engineering Third Edition Octave Levenspiel, page no 94

Reaction equation:



This is main reaction occurs in reactor. As we are using the reaction kinetics to find out the volume of reactor so in the literature we find above reaction is second order. So rate equation for this order is:

$$-r_A = kC_A C_B$$

$$-r_A = kC_A^2$$

$$-r_A = KC_{A0}^2(1-X_A)^2$$

$$X_A = 1 - C_A/C_{A0}$$

We will find the rate constant k by Arrhenius' law:

$$K = k_0 e^{(-E/RT)}$$

Chemical Reaction Engineering Third Edition Octave Levenspiel, page no 27

$$\text{Conversion} = X_A = 97\% = 0.97$$

$$\text{Order of reaction} = n = 2$$

$$C_{A0} = 0.022 \text{ mol/L}$$

By putting the values in respective equation:

$$-r_A = k (0.022)^2 (1-0.97)^2$$

$$-r_A = k (4.356 \times 10^{-7}) \text{ mol}^2/\text{L}^2$$

Here

$$\text{Frequency factor} = k_0 = 1.6 \times 10^9 \text{ L/mol} \cdot \text{Min}$$

$$\text{Activation energy} = E = 19600 \text{ cal/mol}$$

$$\text{Rate law constant} = R = 1.987 \text{ cal/mol} \cdot \text{K}$$

Temperature= $T=258^{\circ}\text{C}=531\text{K}$

So by putting values

$$r_A = 5.95 \times 10^{-6} \text{ mol/L.min}$$

$$F_{A0} = 24 \text{ mol/L}, \quad F_{B0} = 26.97 \text{ mol/L} \quad (\text{from mass balance})$$

$$F_{A0} (\text{average}) = 25.48 \text{ mol/L}$$

Now putting the values in design equation

$$V = 69244.95\text{L}$$

$$V = 69.24 \text{ m}^3$$

$$\text{Volume used} = 69.24/2 = 34.62 \text{ m}^3$$

$$\text{Volume of ellipsoidal head} = 4.67 \text{ m}^3$$

$$\text{Volume of reactor} = 34.6 - 4.67 = 29.93 \text{ m}^3$$

Volumetric flow rate calculation:

	BHET	PTA	EG
mass flow rate (from material balance) in kg/hr	5511.62	4001.908	2990.23
density (g/cm ³)	1.3	1.52	1.11
vol. flow rate (m ³ /hr)	4.23	2.63	2.69

$$\text{Volumetric flow rate} = 8.55 \text{ m}^3/\text{hr}$$

$$\text{For reactor space time} = \tau = V/v_0$$

$$= 29.93/8.55$$

$$\tau = 3.13 \text{ hr}$$

$$\text{Space velocity} = 1/\text{space time}$$

$$= 1/3.13 = 0.268 \text{ hr}^{-1}$$

Length and diameter of reactor:

For low to moderate viscosity, vertical flat end vessel has $L/D = 1.5-2.0$

$$L/D = 2$$

$$L = 2D$$

$$\text{So } V = (\pi/4) D^2 L$$

$$V = (\pi/4) D^2 (2D)$$

$$V = (3.14/2) D^3$$

$$D = (2V/3.14)^{1/3}$$

$$\text{Tank diameter} = D = 2.671 \text{ m}$$

Length of reactor:

$$\text{Length} = L = 5.34 \text{ m}$$

Impeller design:

Impeller agitators are divided into two classes:

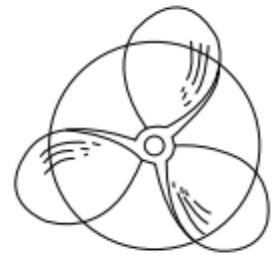
1. Axial flow impeller
2. Radial flow impeller

Axial flow impeller:

The impeller that generates currents parallel to the shaft of the impeller are called axial flow impeller.

Radial flow impeller:

The impeller that generates currents in radial or tangential direction are called radial flow impeller.



There are three main types of impeller for low to moderate viscosity liquids:

- Propeller
- High efficiency impeller
- turbine

As the system of this process is of low to moderate viscosity, so we will have to choose from the categories described above.

Propeller:

A propeller is an axial flow, high speed impeller for liquids of low viscosity. Small impeller turn at full motor speed, either, 1150 or 1750 rpm, greater once turn at 400 to 800 rpm. The direction of rotation is usually chosen to force the liquid downward, and the flow currents leaving the impeller continue until deflected by the floor of the vessel. As described above these are recommended for low viscosity liquids, but are not recommended for gas dispersion in liquid. Also note that they are only axial flow.

High efficiency impeller:

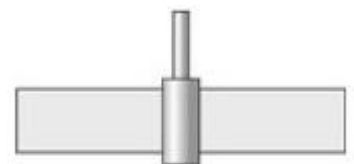
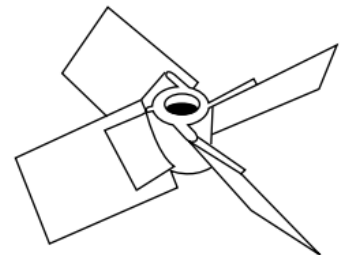
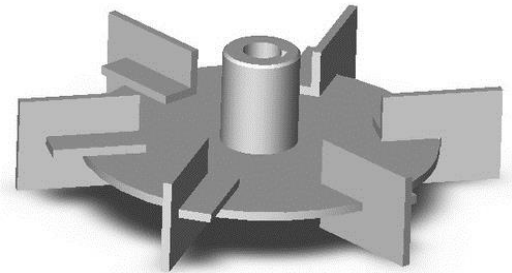
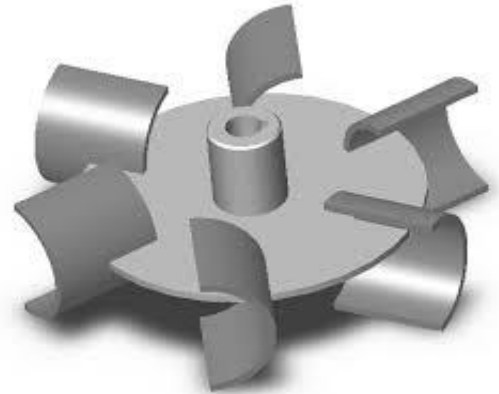
It provide uniform axial flow, better mixing and reduce the power for given flow rate. These impellers are used to mix low to moderate viscosity liquids , but they are not suggested for very viscous liquids or for dispersing gases.

Turbines:

These push the liquid radially and tangentially. The currents they generate travel towards the vessel wall and then flow either upward or downward. In process vessel they typically turn at 20 to 150 rpm. Its type are given by

- **Straight blade turbine impeller** push the liquid radially and tangentially with almost no vertical motion at the impeller. The currents it generates travels outward to the vessel wall and then flow either upward or downward. Such impellers are sometimes called paddles.
- **Concave blade CD 6** is also used for dispersing gases. The 6-blade disk-style concave blade impellers (CBI) are used extensively and economically for gas dispersion in large vessels (in fermenters up to 100,000 gal) at high gas flow rates. This will handle up to 200% more gas without flooding than will the 6BD, and the gassed power draw at flooding drops only about 30%, whereas with a 6 blade, the drop in power draw exceeds 50%.
- The **6-blade disk impeller** is ancient; nevertheless, it still has no peer for some applications. It invests the highest proportion of its power as shear of all the turbine impellers, except those specifically designed to create stable emulsions. It is still the preferred impeller for gas-liquid dispersion for small vessels at low gas rates, it is still used extensively for liquid-liquid dispersions, and it is the only logical choice for use with fast competitive chemical reactions, as will be explained in a later section of this chapter. Lower N_{Re} limit: ~ 5 .

Disc turbine with multiple straight blades mounted on a horizontal disk creates high shear. It is especially useful for dispersing a gas in liquid because at moderate speed the gas is forced to flow radially to the blade tip where it is dispersed by the high shear.



Simple straight-blade turbine

- The 4-blade 45° pitched blade (4 blade) impeller is the preferred choice where axial flow is desired and where there is a need for a proper balance between flow and shear. It is the preferred impeller for liquid-liquid dispersions and for gas dispersion

from the vessel headspace, in conjunction with a lower 6BD or a concave blade disk impeller. Lower N_{Re} limit: ~ 20 .

Impeller selection:

Pitched blade turbine impeller is selected because of

- It is used when good overall circulation is required, because it provide axial flow in addition to radial flow.
- Provide uniform axial flow.
- Used for better mixing.
- It is economical.
- Requires low power to operate.

Unit operations of chemical engineering, McCabe Smith, 7th edition, page no 246

Chemical process equipment selection and design, 3rd edition by J.R.Couper, W.R.Penny, J.P.Fair, S.M.Walas, page no 279

Standard impeller dimensions:

The number of baffles is usually 4, the number of impeller blades ranges from 4 to 16 but is generally 6 or 8. So

Baffles:

Number of baffles selected is 4.

Blades:

Number of blade selected is 4.

Impeller diameter:

$$\frac{\text{impeller diameter}}{\text{tank diameter}} = \frac{D_a}{D} = \frac{1}{3}$$

$$\frac{D_a}{2.67} = \frac{1}{3}$$

$$D_a = 0.89 \text{ m}$$



Pitched blade turbine

Width of impeller blade:

$$\frac{\text{width of impeller blade}}{\text{impeller diameter}} = \frac{W}{D_a} = \frac{1}{5}$$

$$\frac{W}{0.89} = \frac{1}{5}$$

$$W=0.178 \text{ m}$$

Length of impeller blade:

$$\frac{\text{length of impeller blade}}{\text{impeller diameter}} = \frac{L}{Da} = \frac{1}{4}$$

$$\frac{L}{0.89} = \frac{1}{4}$$

$$L=0.2225 \text{ m}$$

Distance of impeller from bottom:

$$\frac{\text{distance of impeller from bottom}}{\text{tank diameter}} = \frac{E}{D} = \frac{1}{3}$$

$$\frac{E}{2.671} = \frac{1}{3}$$

$$E=0.89 \text{ m}$$

Liquid depth:

$$\frac{\text{liquid depth}}{\text{impeller diameter}} = \frac{H}{Da} = 1$$

$$\frac{H}{0.89} = 1$$

$$H= 0.89\text{m}$$

Width of baffle:

$$\frac{\text{width of baffle}}{\text{tank diameter}} = \frac{J}{D} = \frac{1}{12}$$

$$\frac{J}{2.671} = \frac{1}{12}$$

$$J= 0.22258 \text{ m}$$

Unit operations of chemical engineering, McCabe Smith, 7th edition, page no 247

Calculation of power consumption:

Speed of impeller=N=85 rpm

For calculation of power N_p , calculate Reynold's number (Re)

$$Re = \frac{(\text{density of mixture})(\text{speed of impeller})(\text{diamter of impeller})}{\text{viscosity of mixture}}$$

Density of mixture = $\rho = 1389 \text{ kg/m}^3$

Diameter of impeller = $D_a = 0.89 \text{ m}$

Viscosity of mixture = $\mu = 2.3 \text{ cp} = 0.023 \text{ g/cm. sec}$

Putting the values in equation we get

$$Re = 7.614 \times 10^5$$

So flow is turbulent as from graph. Here $Re > 10,000$, $N_p = K_T$

For 4 pitched blade turbine impeller = $K_T = N_p = 1.27$

$$\text{Power} = K_T n^3 D_a^5 \rho$$

$$\text{Power} = 3.75 \text{ hp}$$

Unit operations of chemical engineering, McCabe Smith, 7th edition, page no 262

Thickness of wall of ellipsoidal head:

Formula for thickness is given by:

$$t = \frac{P_i D_i}{2SE - 0.2P_i}$$

For ellipsoidal head, ratio of major to minor axis is 2:1

t = thickness of wall in mm

P_i = maximum allowable design pressure = 0.63 N/mm^2

D_i = internal diameter = 2990 mm

S = joint efficiency = 1.0

E = design stress = 115 N/mm^2

$$t = 8.19 \text{ mm}$$

Thickness of reactor wall:

Formula used for this is:

$$t = \frac{P_i D_i}{2SE - P_i} = 8.2 \text{ mm}$$

Specification sheet:

Equipment name	Continuous Stirred Tank Reactor (CSTR)
Operating temperature	531 K
Volume of reactor	29.93 m ³
Head type	Ellipsoidal
Volume of head	4.67 m ³
Residence time	3.13 hr
Space velocity	0.268 hr ⁻¹
Diameter of reactor / tank	2.671 m
Length of reactor	5.34 m
Impeller type	pitched turbine
Impeller diameter	0.89 m
Width of impeller blade	0.178 m
Number of blades	4
Number of baffles	4
Length of impeller blade	0.2225 m
Distance of impeller from bottom	0.89 m
Depth of liquid	0.89 m
Width of baffle	0.2225 m
Impeller speed	85 rpm
Reynold's number	7.6×10^5
Power required for impeller	3.75 hp
Thickness of head	8.19 mm
Thickness of reactor wall	8.2 mm

7.3. DESIGN OF SHELL AND TUBE HEAT EXCHANGER

FUNCTION

The function of the shell and tube HX to be used here is to add some heat to achieve the desired reaction temperature in endothermic reactions.

SELECTION

The initial and important step is to select the type of HX, which suits our process conditions, keeping in mind the advantages and disadvantages of the following main types of condensers.

1. Double pipe
2. Plate and frame type
3. Shell and tube

Double Pipe HX:

Although, double pipe HX can be cheaply and easily assembled in any pipe fitting shop from standard parts but is not selected due to following reasons.

- a) A small amount of heat transfer surface contained in a single hair pin and so large number of hair pin are needed to accomplish the job and thus required a considerable ground area.
- b) It initials a large number of points where leakage may occur.
- c) Cleaning is time consuming and expensive.

Plates Heat Exchangers

Although these types of heat exchangers are being used mostly in the modern industries because of their adjustably, easy cleaning and more heat transfer at low Reynolds's number, besides their limitation of temperature and pressure. It is not selected because the stream flow rates are very large which is not suited to such types of condensers. Because with large flow rates, the velocity range exceed the allowed limits.

This, we have selected shell and tube heat exchanger to be used for our purpose since it can be handling large flow rate quite well.

Various types of shell and tube heat exchanger are available. The main types are;

Fixed Tube Sheet HX

In fixed tube, the tube sheet is inserted into the shell forming channels, which are integral parts of the shell. In a 1-2 HX, a single channel is employed with a partition to permit the entry and exist of the tube for the same channel. At the opposite and of the

condenser, a bonnet is provided to permit the tube fluid to cross from first pass to the 2nd zone.

DISADVANTAGES

- a) With all fixed tube sheet HX, the outside of the tubes is inaccessible for inspection or mechanical cleaning.
- b) For cleaning inside of the tube, channel cover has to be removed.
- c) Expansion problems are extremely critical, since both passes as well as the shell itself tend to expand differently and cause stress on the stationary tube sheets.

ii. Removable Bundle Condenser

a) Pull-through Floating Head

It consists of a stationary tube sheet, which is clamped between the single channel flange and a shell flange. At the opposite end of the bundle the tubes are expanded into a freely riding floating sheet or floating head. A floating head cover is bolted to the tube sheet and the entire bundle can be withdrawn from the channel end. A shell bonnet closes the shell.

DISADVANTAGES

- i. The number of tubes is reduced.
- ii. An undesirable flow channels between the bundle and the shell is provided.

b. Split Ring Floating Head

Although it is clamped between the floating head cover and a clamp ring placed in back of the tube sheet, which is split in half to permit dismantling.

c. Packed Floating Head

This condenser has an extension on the floating tube sheet, which is confined by means of packing gland. Although entirely satisfactory for shell up to 36-inch ID, the larger packing glands are not recommended for high pressure or services causing vibration.

iii. U-Bend Condenser

It consists of tubes, which are bent in the form of a U and rolled into the tube sheet. The smallest diameter U-bends that can be turned without deforming the outside diameter of the tube at the bend has a diameter of three to four times the outside diameter of the tubing. So, it usually be necessary to omit some tubes at the center of the bundle, depending upon the layout.

MAIN PARTS OF A SHELL & TUBE HEAT CONDENSER

1. TUBES

Heat exchanger tubes are available in a variety of metals, which include steel, copper, admiralty, mint metal, brass, 70-30 copper nickel, aluminum brazen, aluminum and stainless steel.

They are available in standard lengths of 8, 12 and 16 ft with normally 3/4" and O>D. The tubes are of different wall thickness defined by Birmingham wire gauge or gauge of the tube.

2. SHEELS

Shells are fabricated from steel pipe with nominal IPS diameters up to 12 inches. Above 12 and including 24 inches, the actual outside diameter and the nominal pipe diameter are same. The standards wall thickness for shells with inside diameter from 12 to 24-inch inclusive 3/8 inch for operating pressure up to psi.

3. BAFFLES

Baffles are employed to induce turbulence for higher heat transfer coefficient. Although their presence increases the pressure drop on the shell side but the advantage of better mixing and increased turbulence more than offset the pressure drop disadvantage. The center-to-center distance between baffles is called baffle pitch or baffle spacing. The baffle spacing is usually not greater than a distance equal to the inside diameter of the shell or closer than a distance equal to one fifth of the inside diameter of the shell. The most common types of baffles are segmental with 25% cut.

4. TUBE PITCH

The shortest center-to-center distance between two adjacent tubes is called pitch.

Pitch range is 1.25 to 1.5 times tube diameter. Tubes are laid out on either of the two patterns.

1. Square pitch
2. Triangular pitch.

A. Square Pitch

- a) Lesser number of the tubes can be accommodated in the same I.D. shell.
- b) The velocity of the fluid undergoes continuous fluctuation because of the constricted area between the adjacent tubes as compared to the flow area between successive rows.
- c) Lesser turbulence and hence smaller heat transfer coefficient are resulted in square tube pitch.

B. Triangular Pitch

- a) Greater number of tubes is resulted from triangular tube pitch.
- b) Greater turbulence is encountered because the fluid flowing between adjacent tubes at high velocity impinges directly on the succeeding row.

- c) High shell side film coefficient (usually 25% greater than for square pitch) is obtained.

5. FLUID ALLOCATION

The following are the main factors to which a designer should be consider before determining which fluid is to be in the tube side and which should be in the shell side.

1. Cleanability

The shell is difficult to clean and requires the cleanest fluid

2. Corrosion

Corrosion or process cleanliness may result in the use of relatively expensive alloys. Therefore, these fluids are placed inside the tubes in order to save the cost of an alloy shell.

3. Pressure

High-pressure shells, because of their diameter are thick walled and expensive, therefore, high-pressure fluids are placed in the tubes.

4. Temperature

High temperature fluids should be inside the tubes. Very high temperatures reduce the stresses in materials and the effect is sameas the high pressure in determining shell thickness. So, safety of personnel may require the additional cost of insulation if high temperature fluid is in the shell.

5. Hazardous or Expensive Fluids

The most hazardous or expensive fluids should be placed on the tightest side of the condenser, which is the tube side of some types of condensers.

6. Flow Rate

A better overall design may be obtained when the smaller quantity of fluid is placed in the shell. This effect may be due to avoiding multi pass construction with consequent loss of temperature drop efficiency or turbulent flow in shell at low Reynolds's number.

7. Pressure Drop

If pressure drop is critical and must be accurately predicted, then placed that inside the tubes as in shell side there may be a deviation due to leakage and clearance.

8. Viscosity

Viscous fluid should be in the shell side, which will give higher heat transfer coefficient.

9. Fouling

Heavier fouling fluids should be in the tube side because you may have a better velocity control (increasing fluid velocity will decrease the fouling). Mechanical cleaning is easier with straight tubes without removing.

OUTLINE OF THE DESIGN PROCEDURE

The major steps for the design of heat condenser are given as:

1. Define the duty: heat transfer rate, fluid flow rates, and temperature.
2. Collect all the fluid physical properties regarding density, viscosity, thermal conductivity, specific heat etc.
3. Calculate the mean temperature difference (T_m)
4. Select a trial value of the overall coefficient of heat transfer, U
5. Calculate the area required for heat transfer from the equation
$$Q = U A \Delta T_m$$
6. Decide the condenser layout.
7. Calculate the individual film coefficient.
8. Calculate the overall coefficient and compare with the trial value. If the calculated value differs significantly from the estimated value, substitute the calculated value for the estimated value and return to step 5.
9. Calculate the condenser pressure drops, if un-satisfactory return to step 6 or 4 in that order of preference.
10. Optimize the design; repeat step 4-10 as necessary to determine the cheapest condenser that will satisfy the duty within the allowable pressure drop. Usually this will be the one with the smallest area.

Design Calculations

Known Data:

Hot fluid = Sanotherm,

Cold fluid = Dowtherm

Inlet temperature of cold stream

$t_1 = 265^\circ\text{C}$

Outlet temperature of cold stream

$t_2 = 296^\circ\text{C}$

Inlet temperature of hot stream

$T_1 = 325^\circ\text{C}$

Outlet temperature of hot stream

$T_2 = 296^\circ\text{C}$

Physical Properties

Component	Density	Viscosity	Thermal Conductivity	Specific Heat
	kg/m^3	Kg/m.sec	Watt/m.K	kJ/kg. K
Sanotherm	800	0.00039	0.093	2.608
Dowtherm	1062	0.00197	0.61316	4.89

Step 1: Heat Duty

$$Q = m C_p \Delta T$$

$$Q = 1,224,000 \text{ kJ/hr}$$

$$Q = 340,000 \text{ Watt}$$

Now this energy must be provided by sensible heat of Sanotherm

Step 3: Overall Co-efficient

From R.K Sinnott book, the typical range for overall co-efficient is given

$$U_D = 700 - 1000 \text{ W/m}^2 \text{ K}$$

So, assumed

$$U_D = 750 \text{ W/m}^2 \text{ K}$$

Step 4: LMTD

$$LMTD = \left(\frac{(T_2 - T_1) - (t_1 - t_2)}{\ln \left(\frac{T_2 - T_1}{t_1 - t_2} \right)} \right)$$

$$LMTD = \frac{(325 - 296) - (296 - 265)}{\ln \frac{(325 - 296)}{(296 - 265)}}$$

$$LMTD = 29.98^\circ \text{C}$$

$$R = \frac{325 - 296}{296 - 265} = 1.2$$

$$S = \frac{296 - 265}{325 - 265} = 0.51$$

From RK Sinnott Figure 12.19

$$F_T = 0.65$$

So, mean temperature difference:

$$\Delta T_m = LMTD \times F_T$$

$$\Delta T_m = 29.98 \times 0.65^\circ\text{C}$$

$$\Delta T_m = 19.48^\circ\text{C}$$

Step 5: Heat Transfer Area

$$Q = U_D A \Delta T_m$$

$$A = \frac{Q}{U_D \times \Delta T_m}$$

$$A = \frac{340,000}{750 \times 19.48}$$

$$A = 23.27 \text{ m}^2$$

Step 6: Layout & Tube Size

$$\text{Tube OD} = 0.02 \text{ m}$$

$$\text{Tube ID} = 0.016 \text{ m}$$

$$\text{Selected tube length} = L = 4.88 \text{ m}$$

Step 7: Number of tubes

$$\text{Area of one tube} = A_t = \pi d_o L$$

$$A_t = 3.142 \times 0.02 \times 4.88$$

$$A_t = 0.306 \text{ m}^2$$

$$\begin{aligned} \text{Number of tubes} &= \frac{A}{A_t} \\ &= \frac{23.27}{0.306} \end{aligned}$$

$$N_t = 76 \text{ Tubes}$$

$$\text{Tube Passes} = 2$$

$$\text{Tubes per pass} = \frac{76}{2} = 38 \text{ Tubes}$$

$$\begin{aligned} \text{Tube Cross Sectional Area} &= \frac{\pi}{4} D_i^2 \\ &= \frac{3.142}{4} (0.016)^2 \\ &= 0.000209 \text{ m}^2 \end{aligned}$$

$$\begin{aligned} \text{Area per pass} = A_p &= 38 \times 0.000209 \\ &= 0.00742 \text{ m}^2 \end{aligned}$$

$$\begin{aligned}\text{Volumetric flow rate} = Q &= \frac{m}{\rho} \\ &= \frac{16233.8 \frac{kg}{hr}}{800 \frac{kg}{m^3} \times 3600 sec}\end{aligned}$$

$$Q = 0.00565 \text{ m}^3/\text{sec}$$

$$\text{Tube side velocity} = u_t = \frac{Q}{A_P}$$

$$u_t = \frac{0.00565}{0.00742}$$

$$u_t = 0.7596 \text{ m/s}$$

Step 8: Bundle & Shell Diameter

For 2 Tube Passes: From R.K Sinnot table 12.4

$$K_1 = 0.249$$

$$n_1 = 2.207$$

$$D_b = d_o \left(\frac{N_t}{K_1} \right)^{1/n_1}$$

$$D_b = 0.02 \left(\frac{76}{0.249} \right)^{1/2.207}$$

$$D_b = 0.262 \text{ m}$$

For Spilt Ring Floating Head Heat exchanger from figure 12.10 from R.K Sinnot.

$$D_s = 0.262 + 0.05$$

$$D_s = 0.312 \text{ m}$$

Step 9: Tube side Heat Transfer Co-efficient

$$Re = \frac{D_i u_t \rho}{\mu}$$

$$Re = \frac{0.016 \times 0.7596 \times 800}{0.00039}$$

$$Re = 35765$$

$$Pr = \frac{C_P \mu}{K}$$

$$Pr = \frac{2.60 \times 1000 \times 0.00039}{0.093}$$

$$\text{Pr} = 10.90$$

From Graph 12.23 of R.K Sinnot

$$J_h = 7 \times 10^{-3}$$

$$\text{Nu} = J_h \text{Re} \text{Pr}^{0.33} (\mu/\mu_w)^{0.14}$$

$$\text{Nu} = 7 \times 10^{-3} \times 35765 \times (10.90)^{0.33}$$

$$\text{Nu} = 516.7$$

$$\text{Nu} = \frac{h_i d_i}{K}$$

$$h_i = \frac{516.7 \times 0.093}{0.016}$$

$$h_i = 3067 \text{ W/m}^2\text{°C}$$

Step 10: Shell Side Heat Transfer Co-efficient

$$A_s = \left(\frac{P_t - d_o}{P_t} \right) D_s L_B$$

$$L_B = D_s / 5$$

$$L_B = \frac{0.312}{5}$$

$$L_B = 0.0624 \text{ m}$$

$$P_t = 1.25 \times d_o$$

$$P_t = 1.25 \times 0.02$$

$$P_t = 0.025 \text{ m}$$

Now;

$$A_s = \left(\frac{0.025 - 0.02}{0.025} \right) (0.312)(0.0624)$$

$$A_s = 0.00389 \text{ m}^2$$

$$d_e = \frac{1.10}{0.002} (P_t^2 - 0.917 d_o^2)$$

$$d_e = \frac{1.10}{0.02} [(0.025)^2 - 0.917(0.02)^2]$$

$$d_e = 0.0142 \text{ m}$$

Shell Side Velocity:

$$Q = \frac{m}{\rho}$$

$$Q = \frac{4509 \frac{kg}{hr}}{1062 \frac{kg}{m^3} \times 3600 sec}$$

$$Q = 0.001179 \text{ m}^3/\text{s}$$

$$u_s = \frac{Q}{A_s}$$

$$u_s = \frac{0.001179}{0.00389}$$

$$u_s = 0.303 \text{ m/s}$$

$$Re = \frac{D_e u_s \rho}{\mu}$$

$$Re = \frac{0.0142 \times 0.303 \times 1062}{0.00101}$$

$$Re = 4563$$

$$Pr = \frac{C_P \mu}{K}$$

$$Pr = \frac{4.89 \times 1000 \times 0.00101}{0.61316}$$

$$Pr = 15.71$$

$$Nu = J_h Re Pr^{0.33} (\mu/\mu_w)^{0.14}$$

For J_h Graph 12.33 of R.K Sinnot.

$$J_h = 4.7 \times 10^{-3}$$

$$(\mu/\mu_w)^{0.14} = 1$$

$$Nu = (5.8 \times 10^{-3}) (4563) (15.71)^{0.33}$$

$$Nu = 54.09$$

$$Nu = \frac{h_o d_e}{K}$$

$$h_o = \frac{54.09 \times 0.613}{0.0142}$$

$$h_o = 2335.23 \text{ W/m}^2\text{°C}$$

Step 11: Overall Co-efficient

$$\frac{1}{U_D} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2K_w} + \frac{d_o}{d_i} \times \frac{1}{h_{id}} + \frac{d_o}{d_i} \times \frac{1}{h_i}$$

As

We have organic vapor at shell side and organic liquid at tube side

So from Table 12.2

$$h_{od} = h_{id} = 5000 \text{ W/m}^2 \cdot ^\circ\text{C}$$

Putting values will result in:

$$\frac{1}{U_D} = \frac{1}{2335.23} + 0.0002 + \frac{0.02 \ln\left(\frac{0.02}{0.016}\right)}{2 \times 388} + \frac{0.02}{0.016} \times 0.0002 + \frac{0.02}{0.016} \times \frac{1}{3067}$$

$$\frac{1}{U_D} = 1.28 \times 10^{-3}$$

$$U_D = 781.25 \text{ Watt/m}^2 \cdot ^\circ\text{C}$$

Step 12: Pressure Drop

Tube Side Pressure Drop:

$$\Delta P_t = N_P \left(8j_F \left(\frac{L}{d_i} \right) \left(\frac{\mu}{\mu_w} \right)^{-0.14} + 2.5 \right) \frac{\rho u_t^2}{2}$$

From Graph 12.24 for $Re = 161249.5865$

$$j_F = 2.6 \times 10^{-3}$$

$$\begin{aligned} \Delta P_t &= 2 \left(8 \times 7 \times 10^{-3} \left(\frac{4.88}{0.0016} \right) (1)^{-0.14} + 2.5 \right) \frac{800 \times 0.7596^2}{2} \\ &= 8.94 \text{ Pa} \end{aligned}$$

$$\Delta P_t = 8.94 \times 10^{-3} \text{ KPa}$$

Bell's Method at Shell Side

Heat Transfer Co-efficient:

$$h_s = h_{oc} F_n F_w F_b F_L$$

h_{oc} = heat transfer Co-efficient

F_n = Correction factor to allow for effect of No. of vertical tubes rows.

F_w = Window effect correction factor.

F_b = By pass Stream Correction Factor.

F_L = Leakage Correction Factor.

- **Ideal Cross Flow Coefficient h_{oc} :**

$$Re = \frac{D_0 u_s \rho}{\mu}$$

$$= \frac{0.03810 \times 0.2612 \times 1000}{0.0009125}$$

$$Re = 10905.9945$$

And $Pr = 6.2206$

From graph 12.23 R. K Sinnott:

$$J_h = 8 \times 10^{-3}$$

$$Nu = j_h Re Pr^{1/3} \left(\frac{\mu}{\mu_w} \right)^{-0.14}$$

$$= 8 \times 10^{-3} \times 10905.9945 \times 6.2206^{0.33} \times (1)^{-0.14}$$

$$Nu = 159.485$$

$$Nu = \frac{h_{oc} d_e}{K}$$

$$h_{oc} = \frac{159.485 \times 0.61316}{0.0271}$$

$$h_{oc} = 3608.480$$

- **Tube Row Correction Factor F_n :**

As, Reynold Number > 1000 Turbulent Flow

So, Topic 12.9.4 RK Sinnott (Page#856)

$$F_n = 0.97$$

- **Window Correction Factor F_w :**

$$H_b = \frac{D_b}{2} - D_s(0.5 - B_c)$$

$$H_b = \frac{0.262}{2} - 0.312(0.5 - 0.25)$$

$$H_b = 0.053 \text{ m}$$

Bundle Cut:

$$\frac{H_b}{D_b} = \frac{0.053}{0.262}$$

$$\frac{H_b}{D_b} = 0.20 = 20 \%$$

From Graph 12.43 RK Sinnott

Ratio of bundle Cross-sectional area in window zone

$$R_a' = 0.13$$

$$N_w = N_t \times R_a'$$

$$N_w = 76 \times 0.13$$

$$\mathbf{N_w = 9.88}$$

Now, Number of tubes in cross flow Zone

$$N_c = N_t - 2N_w$$

$$N_c = 76 - 2(9.88)$$

$$\mathbf{N_c = 56.24}$$

$$R_w = 2 \left(\frac{N_w}{N_t} \right)$$

$$R_w = 2 \left(\frac{9.88}{76} \right)$$

$$\mathbf{R_w = 0.26}$$

From Graph 12.35

$$\mathbf{F_w = 1.1}$$

- **By-Pass Correction Factor F_b :**

$$A_b = L_b (D_s - D_b)$$

$$L_b = \frac{D_s}{5}$$

$$L_b = \frac{0.312}{5}$$

$$L_b = 0.116 \text{ m}$$

Now:

$$A_b = 0.116 (0.312 - 0.262)$$

$$A_b = 0.0058 \text{ m}^2$$

$$\frac{A_b}{A_s} = \frac{0.0058}{0.00723}$$

$$\frac{\mathbf{Ab}}{\mathbf{As}} = \mathbf{0.802}$$

$$F_b = \exp \left[-\alpha \left(\frac{A_b}{A_s} \right) \left(1 - \left(\frac{2N_s}{N_{cv}} \right)^{0.33} \right) \right]$$

For $Re > 100$

$\alpha = 1.35$

Page# 858 (R.K. Sinnott)

If Sealing Strip is used:

$N_s = 2$

N_{cv} Calculations:

$$N_{cv} = \frac{D_b - 2H_b}{P_t'}$$

Tube vertical pitch = $P_t' = 0.87 \times P_t$

$$= 0.87 \times 0.025$$

$P_t' = 0.02175 \text{ m}$

Baffle cut height = $H_c = 0.25 \times d_s$

$$= 0.25 \times 0.312$$

$$H_c = 0.078 \text{ m}$$

And

$$H_b = 0.1124 \text{ m}$$

$$N_{cv} = \frac{0.312 - 2 \times 0.1124}{0.02175}$$

$N_{cv} = 7.17$

$$F_b = \exp \left[-1.35(0.802) \left(1 - \left(\frac{2 \times 2}{7.17} \right)^{0.33} \right) \right]$$

$F_b = 0.82$

- **Leakage Correction Factor F_L :**

Tube to baffle cut = $C_t = 0.8 \text{ mm}$

Baffle to shell cut = $C_s = 4.8 \text{ mm}$

$$\begin{aligned} A_{tb} &= \frac{C_t \pi d_o}{2} (N_t - N_w) \\ &= \frac{0.8 \times 10^{-3} \times 3.142 \times 0.02}{2} (76 - 9.88) \end{aligned}$$

$A_{tb} = 0.00167 \text{ m}^2$

$$A_{sb} = \frac{C_s d_s}{2} (2\pi - \theta_b)$$

For 20% baffle cut graph 12.43 is used:

$$\Theta_b = 1.95 \text{ rad}$$

$$A_{sb} = \frac{4.8 \times 10^{-3} \times 0.312}{2} (2 \times 3.14 - 1.95)$$

$$A_{sb} = 0.00324 \text{ m}^2$$

$$A_L = A_{sb} + A_{tb}$$

$$A_L = 0.00491 \text{ m}^2$$

$$A_L/A_s = \frac{0.00491}{0.00723}$$

$$= 0.67$$

From Graph 12.35

$$\beta_L = 0.37$$

$$F_L = 1 - \beta_L \left(\frac{A_{tb} + 2A_{sb}}{A_L} \right)$$

$$= 1 - 0.37 \left(\frac{0.00167 + 2 \times 0.00324}{0.00491} \right)$$

$$F_L = 0.38$$

Shell Side Heat Transfer Coefficient:

$$h_s = h_{oc} F_n F_w F_b F_L$$

$$= 5014 \times 0.97 \times 1.1 \times 0.82 \times 0.38$$

$$h_s = 1667 \text{ W/m}^2\text{C}$$

Pressure Drop Calculations

$$u_s = \frac{m}{A \times \rho}$$

$$= \frac{12117.2112}{0.01288 \times 1000}$$

$$u_s = 940.77 \text{ m/hr} = 0.261 \text{ m/s}$$

1. For Cross Flow Zone:

$$\Delta P_C = \Delta P_i F'_B F'_L$$

Where,

$$\Delta P_i = 8 j_F N_{cv} \frac{\rho u_s^2}{2} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

For Re = 7758.9977 and $P_t = 1.25\Delta$ from graph 12.38 (R.K. Sinnott)

$$j_F = 7.5 \times 10^{-2}$$

$$\Delta P_i = 8 \times 7.5 \times 10^{-2} \times 6.857 \times \frac{1000 \times 0.261^2}{2} (1)^{0.14}$$

$$\Delta P_i = 130.789 \text{ N/m}^2$$

$$F_B' = \exp \left[-\alpha \left(\frac{A_b}{A_s} \right) \left(1 - \left(\frac{2N_s}{N_{cv}} \right)^{0.33} \right) \right]$$

For Re > 1000, $\alpha = 4.0$

$$F_B' = \exp \left[-4(0.512) \left(1 - \left(\frac{2 \times 1}{6.857} \right)^{0.33} \right) \right]$$

$$F_B' = 0.5044$$

And $\beta_L' = 0.33$ from Graph 12.40

$$\begin{aligned} F_L' &= 1 - \beta_L' \left(\frac{A_{tb} + 2A_{sb}}{A_L} \right) \\ &= 1 - 0.33 \left(\frac{0.00305 + 2 \times 0.00705}{0.0101} \right) \end{aligned}$$

$$F_L' = 0.4396$$

Now, putting values in equation:

$$\Delta P_c = 130.789 \times 0.5044 \times 0.4369$$

$$\Delta P_c = 29.004 \text{ N/m}^2$$

2. For Window Zone:

From graph 12.43 for 25% baffle cut,

$$R_a = 0.19$$

$$A_w = \left(\frac{\pi D_s^2 R_a}{4} \right) - \left(\frac{N_w \pi d_o^2}{4} \right)$$

$$A_w = \left(\frac{3.142 \times 0.5676^2 \times 0.19}{4} \right) - \left(\frac{12.16 \times 3.142 \times 0.03810^2}{4} \right)$$

$$A_w = 0.0341 \text{ m}^2$$

$$\begin{aligned} u_w &= \frac{w_s}{\rho A_w} \\ &= \frac{12117.2112}{1000 \times 0.0341} \end{aligned}$$

$$= 355.34 \text{ m/hr}$$

$$u_w = 0.0987 \text{ m/sec}$$

as calculated $u_s = 0.261 \text{ m/sec}$

$$u_z = \sqrt{u_s u_w}$$

$$= \sqrt{0.261 \times 0.0987}$$

$$u_z = 0.1605 \text{ m/sec}$$

$$N_{wv} = \frac{H_b}{P'_t}$$

$$= \frac{0.1124}{0.04143}$$

$$N_{wv} = 2.7130$$

$$\Delta P_w = F'_L (2 + 0.6 N_{wv}) \frac{\rho u_z^2}{2}$$

$$= 0.4396 (2 + 0.6 \times 2.7130) \times \frac{1000 \times 0.1605^2}{2}$$

$$\Delta P_w = 18.706 \text{ N/m}^2$$

3. For End Zone:

$$\Delta P_e = \Delta P_i \left(\frac{N_{wv} + N_{cv}}{N_{cv}} \right) F'_b$$

$$\Delta P_e = 130.789 \left(\frac{2.7130 + 6.857}{6.857} \right) 0.5044$$

$$\Delta P_e = 92.071 \text{ N/m}^2$$

4. Total Shell Side Pressure Drop:

$$\Delta P_s = 2 \Delta P_e + \Delta P_c (N_b - 1) + N_b \Delta P_w$$

$$\text{Where, } N_b = \frac{L}{L_b} - 1$$

$$N_b = \frac{1.83}{0.1135} - 1$$

$$N_b = 15.123$$

So, by putting values in upper formula:

$$\Delta P_s = 2 \times 92.071 + 29.004 (15.123 - 1) + 15.123 \times 18.706$$

$$\Delta P_s = 876.656 \text{ N/m}^2$$

Identification:	
Item: Heat Exchanger	
Type: Shell and Tube type	
Function: To increase temperature of Dowtherm	
Heat Duty = 340,000 Watt	
LMTD = 29.98 °C	
Area = 23.27 m ²	
Shell Side	Tube Side
Fluid Handled = Dowtherm vapors	Fluid Handled = Sanotherm
Flow rate = 4509.78 kg/hr	Flow rate = 16233.8 kg/hr
D _s = 0.312 m	No. of Tubes = 76
D _b = 0.262 m	Tube ID = 0.016 m, Tube OD = 0.02 m
u _s = 0.303 m/s	u _t = 0.7596 m/s
Passes = 1	Passes = 2
Pressure Drop = 876.656 N/m ²	Pressure Drop = 894 N/m ²
h _s = 1667.36 W/m ² °C	h _i = 3067.67 W/m ² °C

7.4. Distillation Column

Distillation Column

Distillation is a liquid-liquid separation process which can be carried in continuous or batch systems. The equipment used for distillation is said to be Distillation column. In distillation column, separation is done on the basis of relative volatility. Heat is added to system which increases the temperature of fluids. The fluids having less boiling point evaporates by leaving the liquid behind having lesser volatility.

Types of Distillation Column

1. Batch Column
2. Continuous Column
 - 3.1. The nature of feed they are processing
 - 3.1.1. Binary Column
 - 3.1.2. Multi component column
 - 3.2. Where extra feed exists
 - 3.2.1. Extractive column
 - 3.2.2. Azeotropic column
 - 3.3. Internal Construction
 - 3.3.1. Tray/Plate type
 - 3.3.1.1. Bubble Cap Tray
 - 3.3.1.2. Sieve Tray
 - 3.3.1.3. Valve Tray
 - 3.3.2. Packed column

Why we did choose Sieve Tray Column?

We did choose sieve tray column because of lesser pressure drop and higher separation efficiency with advantage of lesser construction cost.

CALCULATIONS

Feed :

Feed Inlet of EG = 14950.23 Kg/hr

Feed Inlet of Water Vapours = 7804.37

Molecular weight of Ethylene Glycol = 62.07

Molecular weight of water = 18

PRODUCT:

TOP:

E.G = 149.50 kg/hr

Amount of water = 7726.32 kg/hr

BOTTOM:

E.G = 14202.7185 kg/hr

Water = 390.21 kg/hr

TEMPERATURE & PRESSURE:

Pressure of column: 1 atm

Top Temperature = 100.97 °C

Bottom Temperature = 104.88 °C

K- Values:

For Ethylene Glycol : M.W = 62.069

Using Antoine equation

$\ln P_{\text{sat}} = A - (B / T + C)$

$P_{\text{sat}} = 1.074$

$K = P_{\text{sat}} / P = 1.060$

For water:

$$P_{\text{sat}} = 1.551$$

$$K = P_{\text{sat}} / P = 1.530$$

Light Key = Ethylene Glycol

Heavy key = water

Relative Volatility of water = 1

Relative Volatility of EG=0.6536

Minimum Reflux Ratio

Using Underwood equation

$$\frac{\alpha_A x_{fA}}{\alpha_A - \theta} + \frac{\alpha_B x_{fB}}{\alpha_B - \theta} = 1 - q$$

As feed is a mixture of vapors and liquid so $q = 0.2857$

By iteration on Micro-soft Excel ; Value of $\theta = 0.872$

Using equation of Minimum reflux ratio,

$$\frac{\alpha_A x_{DA}}{\alpha_A - \theta} + \frac{\alpha_B x_{DB}}{\alpha_B - \theta} = R_m + 1$$

$$R_{\text{min}} = 1.88$$

For optimum Reflux Ratio:

Using Rule Of Thumb

$$R = 2.82$$

Calculating Minimum No. Of Stages, Using Fenske's equation

$$N_{\min} = \frac{\ln[(x_{LK}/x_{HK})_D (x_{HK}/x_{LK})_B]}{\ln(\alpha_{LK/HK})_{\text{average}}}$$

$$= 9.42375$$

Theoretical No. of Plates:

$$N - N_{\min} / N + 1 = 0.75 \left[1 - \left(R - R_{\min} / R + 1 \right)^{0.566} \right]$$

$$N = 28$$

Column Efficiency:

Average temperature of column = 375.925°C

Avg viscosity = 0.792832

$$\alpha_{\text{ave.}} = 1$$

$$E_o = 51 - 32.5 \left[\log \left(\mu_{\text{avg}} \cdot \alpha_{\text{avg}} \right) \right]$$

$$= 54.276\%$$

So, **No. of actual trays = 31**

8.3.3.5. Location of Feed Plate:

$$\log \left(\frac{N_D}{N_B} \right) = .206 \log \left[\left(\frac{B}{D} \right) \left(\frac{x_{HK}}{x_{LK}} \right) \left(\frac{(x_{LK})_B}{(x_{HK})_D} \right)^2 \right]$$

$$N - N_B / N_B = 0.1947$$

$$0.1947 N_B = N - N_B$$

$$N = 1.1947 N_B$$

$$N_B = 26$$

Material Balances:

Above feed plate (Rectifying section):

$$L_n = (D)(R_{min})$$

$$L_n = 14672.2156 \text{ kg/hr}$$

$$V_n = L_n + D$$

$$V_n = 22476.5856 \text{ Kg/hr}$$

Bottom Calculations:

Below feed plate (Stripping section);

$$L_m = L_n + F$$

$$L_m = 48222.3156 \text{ kg/hr}$$

$$V_m = L_m - B$$

$$V_m = 33271.1656 \text{ kg/hr}$$

Determination of the Column Diameter

Vapor load at Top: (light key)

$$Q_v = \frac{V_m}{\rho_v \times 3600} = 3.36 \text{ m}^3/\text{Sec} \quad [V_m = V_n]$$

Liquid load at Top

$$Q_L = \frac{L_m}{\rho_L \times 3600} = 0.01217 \text{ m}^3/\text{Sec} \quad [L_m = L_n]$$

Vapor load at bottom

$$Q_v = \frac{V_m}{\rho_v \times 3600} = 4.96 \text{ m}^3/\text{Sec}$$

Liquid load at bottom

$$Q_L = \frac{L_m}{\rho_L \times 3600} = 0.01327 \text{ m}^3/\text{Sec}$$

Spacing between the plates = 2 ft = 0.61 m [table 15-5: Plant Design by Peters]

Tray Dynamics:

i) Flow Parameter:

$$F_{LV} = \left(\frac{L_m}{V_m} \right) \left(\frac{\rho_v}{\rho_L} \right)^{0.5} = 0.0723$$

Capacity Parameter:

Net vapor velocity at flooding;

$$V_{nf} = C_{sb} \left(\frac{\sigma}{20} \right)^{0.2} \left(\frac{\rho^l - \rho^v}{\rho^v} \right)^{0.5}$$

$$= 2.124 \text{ m/sec}$$

Assume 80% of flooding;

$$V_N = 0.80 V_{nf}$$

$$= 1.7 \text{ m/sec}$$

$$\text{Net-Area} = A_n = \text{Volumetric flow rate of vapor} / \text{Actual vapor velocity}$$

$$= 3.5 \text{ m}^2$$

$$A_c = A_n + 0.15 A_c$$

$$= 4.177 \text{ m}^2$$

$$\text{Diameter of Column} = A_c = \frac{\pi}{4} D^2$$

$$\Rightarrow D = \sqrt{\frac{4 \times A_c}{\pi}}$$

$$D = 2.306 \text{ m}$$

Flooding Check

$$V_n = \frac{Q_V}{A_n} = 0.946 \text{ m/Sec}$$

$$\text{Now } F = (V_n/V_{nf}) \times 100$$

$$F = 45\%$$

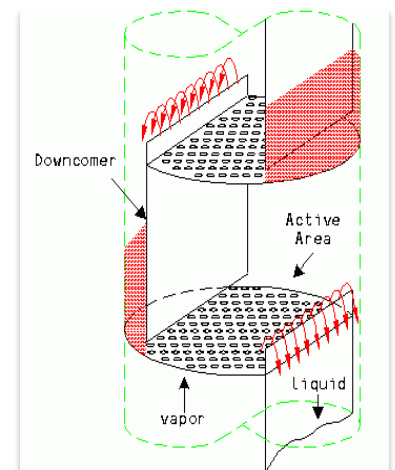
Calculation of Entrainment

$$\text{As FLV} = 0.0723 \text{ and } F = 45\%$$

From Figure 11.29: Richardson and coulson volume 6, we calculate

$$\psi = 0.007$$

ψ = Fractional Entrainment factor



$$\text{Net area} = A_n = 3.5509 \text{ m}^2$$

$$\text{Tower Cross-sectional Area} = A_c = 4.177 \text{ m}^2$$

$$\text{Active area} = A_a = A_c - 2A_d$$

$$A_a = 2.937 \text{ m}^2$$

$$\text{Down comer area} = A_d = 0.15 A_c = 0.62 \text{ m}^2$$

$$\text{Hole area} = A_h = 0.05 A_a = 0.14685 \text{ m}^2$$

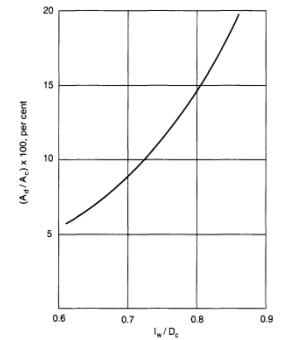
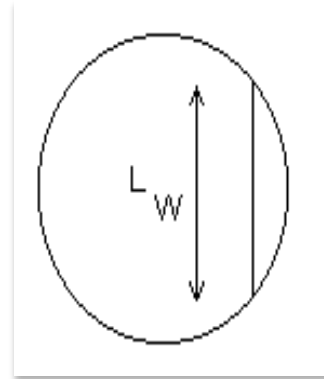


Figure 11.31. Relation between downcomer area and weir length

Weir Length (L_w):

$$A_d/A_c = 14.86\%$$

From fig. 11.31 (pg 573) RC-6

$$L_w/D_c = 0.80$$

$$L_w = (2.306)(0.80)$$

$$\text{Length of weir } (L_w) = 1.84 \text{ m}$$

Estimation Of Weep point:

The minimum vapor velocity for weeping is estimated by;

$$\bar{U}_{h(\min)} = \frac{K_2 - [0.90 - (25.4 - d_h)]}{(\rho_v)^{0.5}}$$

K_2 = Constant depending upon the depth of the clear liquid

$$\bar{U}_{h(\min)} = 19.90 \text{ m/sec}$$

$$\text{Actual Min. Vapor Velocity} = 52.34 \text{ m/sec}$$

So, Minimum vapor state rate will be below the weeping point.

Tray Pressure Drop

$$H_t = H_d + (H_w + H_{ow}) + H_r = 133.56 \text{ mm}$$

Total Pressure Drop:

$$\Delta P_t = (9.81 \times 10^{-3}) H_t \times \rho_L = 1441.241 \text{ Pa}$$

$$= 0.2089 \text{ Psi /per tray}$$

Height of Distillation Column

$$\text{Height of column } H_c = (N_{\text{act}} - 1) H_s + \Delta H$$

$$\text{Total height of column} = 20.37 \text{ m}$$

The Length/Diameter ratio of a tower should be no more than 30 and preferably below 20

$$\text{Length/Diameter} = 20.374 / 2.306 = 8.83$$

SPECIFICATION SHEET

REFLUX RATIO	1.88	WEIR LENGTH	1.84 m
NO. OF STAGES	28	HOLE AREA	0.14685 m ²
PRESSURE	1 atm	TRAY SPACING	0.61 m
HEIGHT OF COLUMN	20.37 m	DOWN COMER AREA	0.62 m ²
DIAMETER OF COLUMN	2.306 m	EFFICIENCY	54.276 %
AREA	4.177 m ²	Minimum Vapor Velocity for Weeping	19.90 m/sec
FLOADING	45 %	PRESSURE DROP	0.2089 psi/per tray
ACTIVE AREA	2.937 m ²	L/D	8.83

7.5. Furnace Design

Furnace:

Furnace is also called fired heater or direct fired heater and it is a heat transfer equipment used in process industry to achieve high temperatures. In our process we are using furnace to heat Sano therm upto 325-328°C which is further used to pre heat dowtherm. Furnace is providing heat required in endothermic reactions in our process. Furnace is a closed air system which attains very high temperature as compare to other open air systems. Furnace design varies according to its function, heat duty, method of air induction and type of fuels.

Working:

Fuel flows in burner and burned with air provided by fans on lower side of furnace. Burners can be at floor, walls or roof. The flame heats the fluid in radiant section (tubes). The fluid passing through radiant section is heated up to desired temperature. Flue gases produced in radiant section are further interacted in convection section where heat of flue gases is recovered.

Sections:

1. Radiant Section
2. Convection Section
3. Radiant Coil
4. Burner
5. Blower
6. Stack
7. Insulation
8. Foundations
9. Access doors

Design Methods:

1. Lobo and Evans
2. Wilson, Lobo and Hottel
3. The Orrok-Hudson Equation
4. Wohlenberg simplified method

Design Calculation:

1. Average Heat Flux

Assumed avg heat flux = 11000 BTU/hr.ft²

$$\frac{\Sigma Q}{\alpha A_{cp}} = 2 \times 11000 \frac{BTU}{hr.ft^2} = 22000 \frac{BTU}{hr.ft^2}$$

$$\frac{\Sigma Q}{\alpha A_{cp}} = 249702.464 \frac{KJ}{hr.m^2}$$

2. Overall exchange factor

Overall exchange factor = f = 0.57 (reference: D. Q Kern Book)

$$\frac{\Sigma Q}{\alpha A_{cp} f} = 438074 \frac{KJ}{hr.m^2}$$

3. Tube Surface Temperature:

Inlet temperature of fluid in tubes = 296°C

Outlet temperature of fluid = 325°C

We can assume tube surface temperature above leaving temperature = 350°C
= 662°F

Evaluated T_G from fig 19.14

T_G = 1580°F = 860°C

4. Total Heat Duty:

Q_{net} = 63,940,000 KJ/hr

Heat Liberated by fuel;

Q_F = Q_{net}/η = 416,725,000/0.75

Q_F = 85,253,333.3 KJ/hr

Fuel consumption = q_F = Q_F/Lowering heating value of fuel

Methane is fuel and lowering heating value of methane = 50,000 KJ/kg

So

q_F = 1705 kg/hr = 106.6 Kmoles/hr

25% excess air corresponding to $17.44 \frac{kg \text{ of air}}{kg \text{ of fuel}}$

$$\text{Air required} = q_F \times \frac{\text{kg of air}}{\text{kg of fuel}} = 1705 \times 17.44$$

$$q_A = 29735.2 \text{ kg/hr}$$

$$\text{Entering temperature of air} = 25^\circ\text{C}$$

$$\text{Enthalpy of entering air} = H_A = 25.072 \text{ KJ/kg}$$

$$Q_A = H_A \times q_A = 29735.2 \times 25.072$$

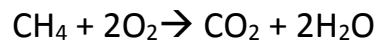
$$Q_A = 745,520 \text{ KJ/hr}$$

$$Q_w = 2\% \text{ of } Q_F \quad (\text{Heat loss through furnace wall})$$

$$Q_w = 1,705,066.67 \text{ KJ/hr}$$

$$Q_{\text{net}} = 84,293,786.33 \text{ KJ/hr}$$

5. QG from reaction equation



According to mass balance

$$\text{Oxygen consumed} = 2 \times 106.6 = 213.2 \text{ Kmol/hr}$$

As Oxygen is supplied in 25% excess

$$\text{Total Oxygen supplied} = 1.25 \times 213.2 = 266.5 \text{ Kmol/hr}$$

$$\text{Oxygen in flue gas} = 266.5 - 213.2 = 53.3 \text{ Kmol/hr}$$

$$\text{Nitrogen in flue gases} = 266.5 \times \frac{0.79}{0.21} = 1002.54 \text{ Kmol/hr}$$

$$\text{CO}_2 \text{ in flue gases} = 106.6 \text{ Kmol/hr}$$

$$\text{H}_2\text{O in flue gases} = 2 \times 106.6 = 213.2 \text{ Kmol/hr}$$

$$H \text{ of } \text{CO}_2 \text{ at } T_G = 40,729 \text{ KJ/Kmol}$$

$$H \text{ of } \text{H}_2\text{O} \text{ at } T_G = 31540.6 \text{ KJ/Kmol}$$

$$H \text{ of } \text{N}_2 \text{ at } T_G = 25857.2 \text{ KJ/Kmol}$$

$$H \text{ of } \text{O}_2 \text{ at } T_G = 27433.8 \text{ KJ/Kmol}$$

$$Q_{CO_2} = 40,729 \text{ KJ/Kmol} \times 106.6 \text{ Kmol/hr} = 4,341,711.4 \text{ KJ/hr}$$

$$Q_{H_2O} = 31540.6 \text{ KJ/Kmol} \times 213.2 \text{ Kmol/hr} = 6,724,455.92 \text{ KJ/hr}$$

$$Q_{N_2} = 25857.2 \text{ KJ/Kmol} \times 1002 \text{ Kmol/hr} = 25,908,914 \text{ KJ/hr}$$

$$Q_{O_2} = 27433.8 \text{ KJ/Kmol} \times 53.3 \text{ Kmol/hr} = 1,462,221.54 \text{ KJ/hr}$$

$$Q_G = Q_{CO_2} + Q_{H_2O} + Q_{N_2} + Q_{O_2}$$

$$Q_G = 38,437,301.6 \text{ KJ/hr}$$

6. Total Q in Radiant Section

$$Q_R = Q_{net} - Q_G$$

$$Q_R = (84,293,786.33 - 38,437,301.6) \text{ KJ/hr}$$

$$Q_R = 45,856,484 \text{ KJ/hr}$$

7. Tube Specifications:

Assume the length $L = 8 \text{ m}$

Inner Diameter of the tube $D = 0.25 \text{ m}$

Outer Diameter of the tube $= 0.285 \text{ m}$

C-C distance $= 0.57 \text{ m}$

Surface area of tubes $= A_s = \pi DL$

$$A_s = 7.15 \text{ m}^2$$

8. Heat Transfer per tube:

$$Q_{tube} = \text{flux} \times A_s$$

$$Q_{tube} = 1,785,369.422 \text{ KJ/hr}$$

9. Number of tubes

$$N = Q/Q_{tube}$$

$$N = 45,856,484 / 1,785,369.422$$

$$N = 26$$

10. Diameter of Furnace:

$$2\pi R = \text{number of tubes} \times \text{C-C distance}$$

$$R = (26 \times 0.57) / 2\pi$$

$$R = 2.35 \text{ m}$$

$$D = 4.7 \text{ m}$$

11. Effectiveness Factor

$$\frac{C - C \text{ distance}}{OD \text{ of tubes}} = 2$$

From fig 19.11

$$\alpha = 0.89$$

12. Cold surface plane area A_{cp}

$$A_{cp} = (\text{No. of tubes}) (\text{Length of each tube}) (\text{C-C distance})$$

$$A_{cp} = (26) (8) (0.57) = 118.56 \text{ m}^2$$

$$A_{cp}/\text{tube} = 4.56 \text{ m}^2$$

$$\alpha A_{cp}/\text{tube} = 4.01 \text{ m}^2$$

13. Total area of furnace

Total Area of Radiant Section

$$A_{rad} = 2 \times \pi \times 2.35 \times 8 + 2 \times \pi \times 2.35^2$$

$$A_{rad} = 152.82 \text{ m}^2$$

Total Area of Stack

$$A_{stack} = 2 \times \pi \times 1.175 \times 4 + 2 \times \pi \times 1.175^2$$

$$A_{stack} = 38.21 \text{ m}^2$$

Total Area of Furnace

$$A_t = 152.82 + 38.21$$

$$A_t = 191.03 \text{ m}^2$$

$$A_R = A_t - \alpha A_{cp} \quad (\text{Effective refractory surface})$$

$$A_R = 191.03 - (0.89 \times 118.56)$$

$$A_R = 85.511 \text{ m}^2$$

$$A_R / \alpha A_{cp} = 0.8195$$

14. Mean Beam Length

Using table 19.1

$$L = 2/3 \times D$$

$$L = 2/3 \times 4.7$$

$$L = 3.13 \text{ m}$$

15. Emmisivity of gas:

Partial pressures of CO₂ and H₂O with 25% excess air are

$$p_{\text{CO}_2} = 0.1084 \quad p_{\text{CO}_2} \times L = 0.3392$$

$$p_{\text{H}_2\text{O}} = 0.1248 \quad p_{\text{H}_2\text{O}} \times L = 0.39$$

From Fig. 19.12 and 19.13, the emissivity of the gas can be evaluated.

$$\epsilon_g = \left[\frac{(q_{\text{CO}_2} \text{ at } P_{\text{CO}_2} L + q_{\text{H}_2\text{O}} \text{ at } P_{\text{H}_2\text{O}} L)_{T_g} - (q_{\text{CO}_2} \text{ at } P_{\text{CO}_2} L + q_{\text{H}_2\text{O}} \text{ at } p_{\text{H}_2\text{O}} L)_{T_s}}{(q_b)_{T_g} - (q_b)_{T_s}} \right]$$

Where

$$T_s = 662^\circ\text{F} \quad \text{and} \quad T_g = 1580^\circ\text{F}$$

And

$$q_b = 0.173 \epsilon_b \left(\frac{T}{100} \right)^4 \quad \text{and} \quad \epsilon_b = 1.00$$

$$q_b \text{ at } T_g = 43555$$

$$q_b \text{ at } T_s = 2741$$

By putting all values in equation we get

$$\epsilon_g = 0.3949$$

Specification Sheet

Equipment Name	Furnace
Operation	Continuous
Shape	Cylindrical
Material of construction	Alloy
Inlet Temperature	286°C
Outlet Temperature	325°C
Flue gases temperature	860°C
Total Heat Duty	85,253,333.3 KJ/hr
Heat Flux	$249702.464 \frac{KJ}{hr.m^2}$
Correction factor assumed	0.57
Correction factor calculated	0.53
No. of tubes	26
Furnace diameter	4.7 m
OD of tubes	0.285 m
ID of tubes	0.25 m
C-C distance	0.57 m
Fuel type	Natural gas
Fuel Required	1705 Kg/hr
Emissivity	0.36

8. Instrumentati on and Process Control

8.1 Introduction to instrumentation:

The overall objective of any material processing plant is to convert raw material to desired product using available sources of energy and utilities in the most economical manner. The ability of instruments to measure the process conditions depends on various features like working principle, cost, installation and maintenance.

The process control is the automatic control of output variable by sensing the parameter under observation from the process and comparing it to the desired or set level and feeding an error signal back to control an input variable. A control loop is established to achieve the purpose of process control.

It is basically an electrical circuit provided with a power supply, actuators and controller. The main element however remains the instrument that measures and indicates the values of variables to be monitored and this measuring element includes the sensor, transducer and transmitter.

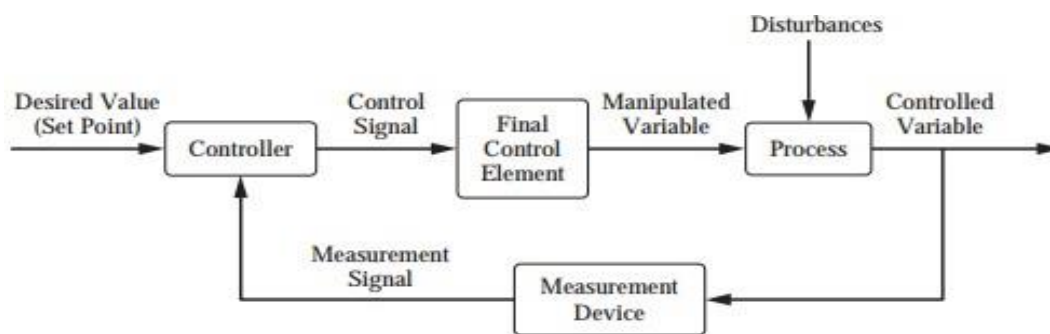


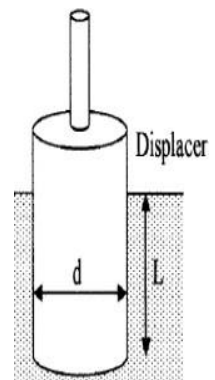
FIGURE 1-1
Generalized process control system.

8.1.1 Instrumentation over esterification tank:

Two level measuring instruments are used for esterification tank.

- 1) Displacer
 - 2) Bubbler tube
- **Displacer:**

The displacer basically follows the Archimedes Principle that when a body is immersed in a fluid it appears less in weight equal to that of the fluid displaced. They are force balance gauges and the buoyant force on the displacer object is proportional to the weight of the fluid displaced and indicate level change. A spring is loaded with weighted displacers and when the buoyancy changes after immersing in water, the spring moves upward transmitting the buoyant force. Thus level is indicated by a force sensor.



$$F = (1/4) \gamma \pi d^2 L$$

Where,

γ is specific weight of the liquid,

d is float diameter;

L is length of the displacer submerged in the liquid.

- **Bubbler tube:**

This instrument essentially consists of a tube with its open end inserted near the bottom of the tank. Inert gas or clean air is forced out of the tube and it escapes as bubbles into the liquid in the tank. The pressure required to force out the liquid is equal to pressure at the end of the tube. The pressure in the tube is equal to hydrostatic head which is measure of the level. As the level varies the pressure in the tube also varies correspondingly.

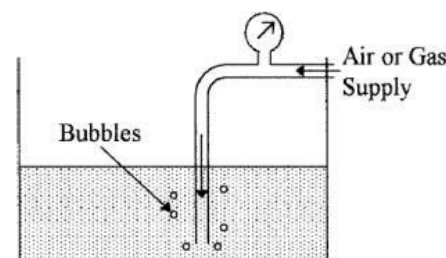


Figure 5.4: Bubbler Tube Technique

The system consists of a pipe, an air supply, a pressure transmitter and a differential pressure regulator. The specific weight of the gas is considered negligible and pressure is equal to the product of depth of liquid and specific weight of liquid.

Advantages:	Disadvantages:
• Suitable for corrosive liquids, slurries or other fluids with entrained solids because coming out of the tube prevent any process fluid to enter the system. • No moving parts • Easy installation • Good accuracy and repeatability	• Not suitable for pressurized tank • Sediments/impurities may block tube • Tank must be freely vented

8.2 Introduction to control:

A chemical plant is an arrangement of processing units (reactors, heat exchangers, distillation column, absorber, evaporators, tanks etc) integrated with another in systematic and rational manner. The plant overall is to convert certain raw materials (input, feedstock) into desired products using available sources of energy, in the most economical way.

During its operation, a chemical plant must satisfy several requirements imposed by its designers and the general technical, economic and social conditions in the presence of ever-changing external influences (disturbances).

8.2.1 Incentives/objectives of process control:

Among such requirements are following:

- Safety
- Production specification
- Environmental regulations
- Operational constraints
- economics
- **Safety:**

The safe operation of a chemical process is a primary requirement for the well being of people in the plant and for its continued contribution to the economic development. Thus, the operating pressure, temperature, concentration of chemical and so on should be in allowable limits.

- **Production specifications:**

A plant should produce the desired amounts and quality of final products. For example, we require the production of 2 million pounds of ethylene per day 99.5% purity. Therefore, a control system is needed to ensure the production level and purity is satisfied.

- **Environmental regulations:**

Various federal and state laws may specify the temperature, concentration of chemicals and flow rates of the effluents from a plant be satisfied through the operation of plant.

- **Operational constraints:**

The various types of equipments used in chemical plant have constraints inherent to their operation. Such constraints should be satisfied through the operation of plant.

- **Economics:**

The operation of plant must confirm with the market conditions, that is, availability of raw material and the demand of the final product. Furthermore, it should be economical as possible in its utilization of raw material, energy, capital and human labor. Thus, it is required that the operating conditions are controlled at given optimum level of minimum operating cost, maximum profit and so on.

8.2.2 Why control system is required?

There are three general classes of needs that a control system is required:

- Suppressing the influence of external disturbances
- Ensuring the stability of a chemical process
- Optimizing the performance of chemical process

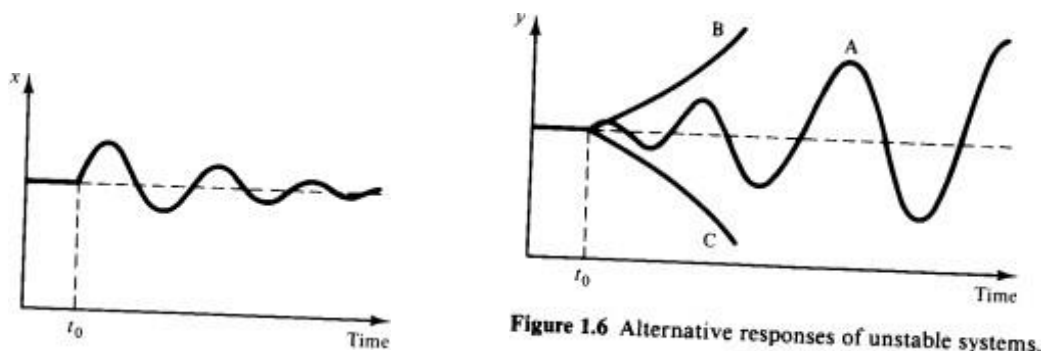
- Suppressing the influence of external disturbances:

Suppressing the influence of external disturbances on process plant is the most common objective of controller in a chemical plant. Such disturbances which denotes the effect that the surrounding have on reactor, distillation column , separator, heat exchanger, compressor and so on are usually out of the reach of the human operator. Consequently we need a control mechanism that will make the proper changes on the process to cancel the negative impact that such disturbances may have on the desired operation of plant.

- Ensuring the stability of a chemical process:

Consider the behavior of variable x as shown in figure. Notice that at time $t=t_0$ the constant value of x is disturbed by some external factors, but that as time progresses the value of x returns to its initial values and stay there. If x is a process variable such as temperature, pressure, concentration or flow rate, we can say that the process is stable or self regulating and no external intervention for its stabilization.

In contrast to the behavior described above the variable y does not return to its initial value after it is disturbed by external influences. Processes whose variables follow the pattern indicate by y is unstable process and require external control for process stabilization.



- Optimizing the performance of chemical process:

Safety and satisfaction of production specification are two principle operational objectives for chemical plant. Once these are achieved, the next goal is how to make the operation of the plant more profitable. This task is under taken by the automatic controllers of the plant and its human operators.

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8.2.3 Classification of variables in chemical process:

The variables (flow rate, temperature, pressure, concentration etc) associated with a chemical process with a chemical process are divided into two groups:

- *Input variables:*

It denotes the effect of surrounding on chemical process.

- Variables whose values can be adjusted freely by human operator or by chemical mechanism.
- Disturbances, if their values are not the result of an adjustment by an operator or a control system.

- *Output variables:*

It denote the effect of process on the surrounding.

- Measured output variables, if their values are known by directly measuring them.
- Unmeasured output variables, if their values cannot be measured directly.

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8.2.4 Hardware elements of control system:

In every control configuration we can distinguish the following hardware elements:

- The chemical process
- Sensor
- Transducer
- Transmission lines
- Controller
- Final control element
- Recording element
- The chemical process:

It represents the material equipment together with the physical or chemical operations that occur there.

- Sensor:

Such instruments are used to measure the disturbances, the control output variables, or secondary output variables are main sources of information about what is going on in the process. For example

Thermocouples or resistance thermometers are used to measure the temperature.

- Transducer:

Many instruments cannot be used for control until they are converted to physical quantities (such as electric voltage, current or pneumatic signal) which can be transmitted easily. Transducers are used for this purpose. For example, strain gauges are metallic conductors

whose electrical resistance changes when they are subjected to mechanical strain. Thus, they can be used to convert a pressure signal to electric one.

- Transmission lines:

These are used to carry the measurement signals from measuring elements to controller. Many signals coming from a measuring device is very weak and cannot be transmitted over a long distance. In such cases transmission lines are equipped with amplifiers which raise the level of signals.

- Controllers:

This is hardware elements that have 'intelligence'. It receive the information from measuring devices and decide what action is to be taken.

- Final control element:

This is the hardware element that implement in real life the decision takes by controller.

- Recording element:

These are used to provide a visual demonstration how a chemical process behaves. Usually, the variables recorded are the variables that are directly as a part of control system.

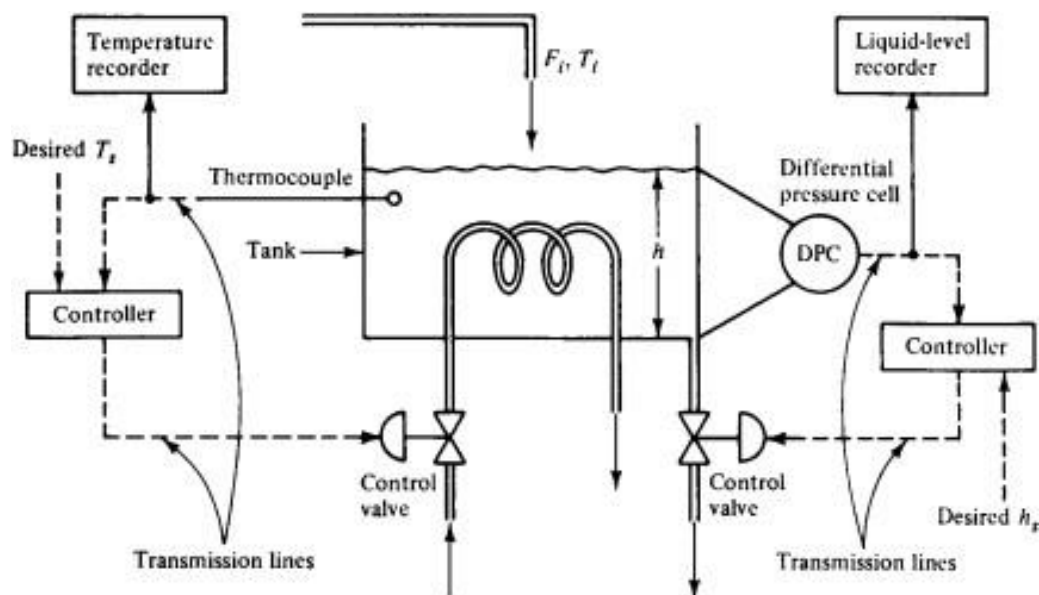


Figure 3.1 Hardware elements for the feedback control of a stirred tank heater.

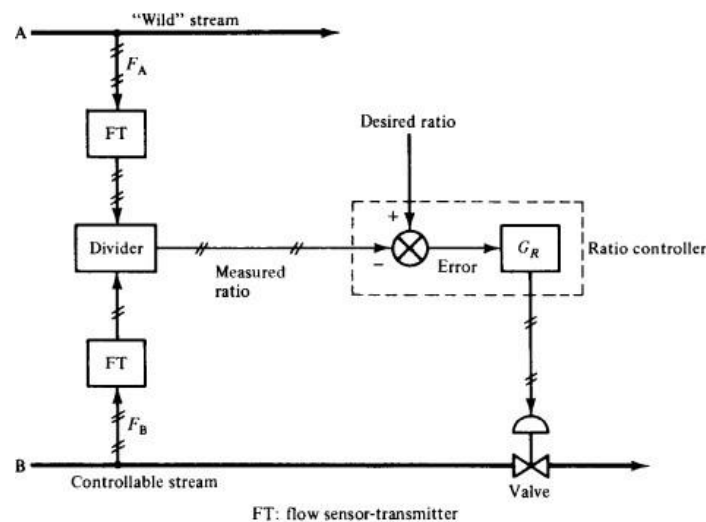
8.2.5 Control over mixer:

Ratio controller configuration is used in mixer to control (E/T) ratio in mixing tank.

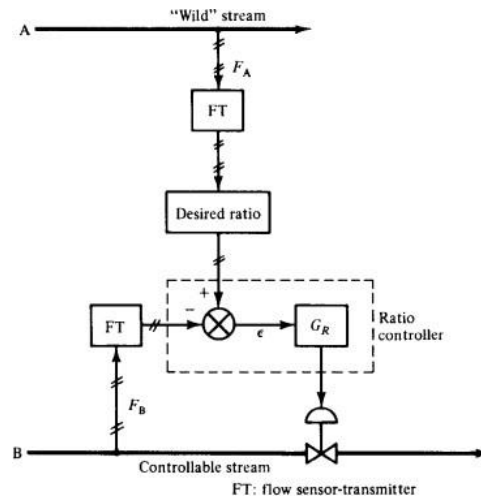
Ratio control is a special type of feed forward control where two disturbances are measured and held in constant ratio with each other. Is mostly used to control the flow rate of two streams. Both flow rates are measured but only one can be controlled. The stream whose flow rate is not under control is called 'wild stream'.

Stream A is a wild stream. It has two configurations.

- In configuration 1 we measure both flow rates and take their ratio. This ratio is compared to desired ratio (set point) and the deviation (error) between the measured and desired ratio can be constitutes the actuating signals for ratio controller.



- In configuration 2 flow rate of wild stream A and multiply it by the desired ratio. The result is the flow rate that B would have and constitutes the set point value which is compared to the measured flow rate of stream B. The deviation constitutes the actuating signals for controller, which appropriately adjust the flow of B.



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8.2.6 Control over esterification tank:

The schemes used for reactor control depend on the process and the type of reactor. If a reliable on-line analyser is available, and the reactor dynamics are suitable, the product composition can be monitored continuously and the reactor conditions and feed flows controlled automatically to maintain the desired product composition and yield. More often, the operator is the final link in the control loop, adjusting the controller set points to maintain the product within specification, based on periodic laboratory analyses.

Reactor temperature will normally be controlled by regulating the flow of the heating or cooling medium. Pressure is usually held constant. Material balance control will be necessary to maintain the correct flow of reactants to the reactor and the flow of products and unreacted materials from the reactor.

A typical reactor control scheme is shown in Figure.

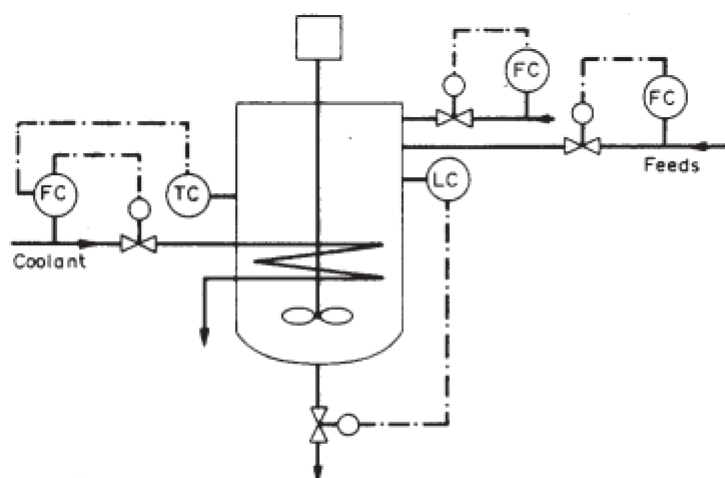


Figure 5.23. A typical stirred tank reactor control scheme, temperature: cascade control, and reagent: flow control

Chemical engineering design Richard coulson, 4th edition,

6th volume, ch # 5, pg # 233

In the following loop, esterification tank is control by temperature parameter. We give a set point as temperature reaches at that set point composition, level and height control automatically. It is feedback control configuration.

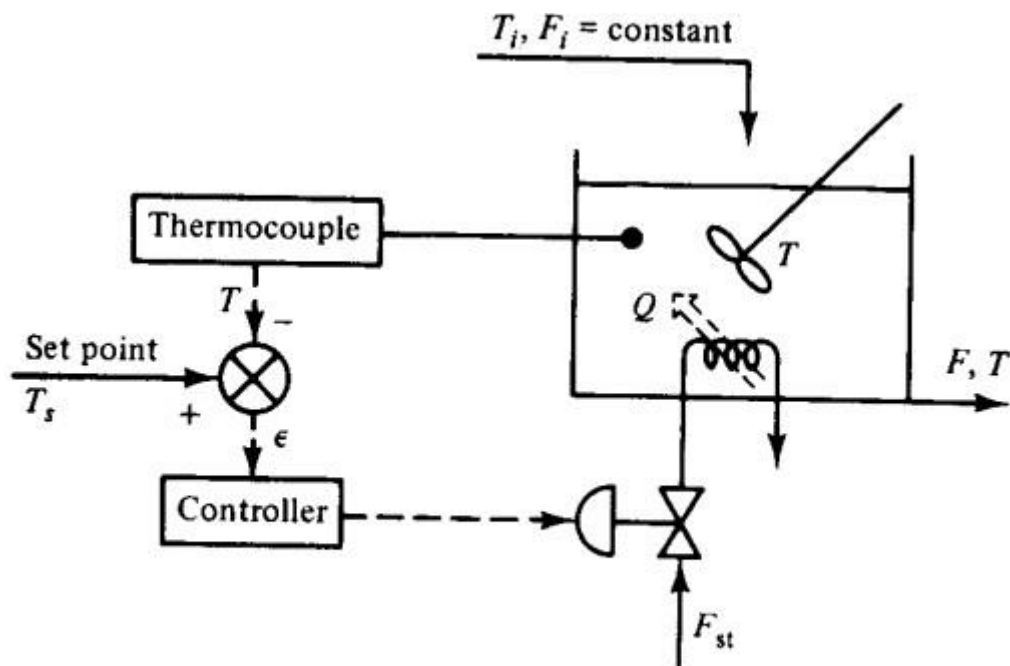


Figure 1.2 Feedback temperature control for a tank heater.

Chemical process control An introduction to theory and practice

by George Stephanopoulos ch # 01 , pg # 5

9.HAZOP Study

9.1 Introduction to HAZOP:

Hazard and Operability Analysis (HAZOP) is a structured and systematic technique for system examination and risk management. The HAZOP technique was initially developed to analyze chemical process systems, but has later been extended to other types of systems and also to complex operations and to software systems. HAZOP is based on a theory that assumes risk events are caused by deviations from design or operating intentions. Identification of such deviations is facilitated by using sets of “guide words” as a systematic list of deviation perspectives. This approach is a unique feature of the HAZOP methodology that helps stimulate the imagination of team members when exploring potential deviations. The HAZOP is a qualitative technique based on guide-words and is carried out by a multidisciplinary team (HAZOP team) during a set of meetings. HAZOP is also commonly used in risk assessments for industrial and environmental health and safety applications.

9.2 HAZOP characteristic:

HAZOP is best suited for assessing hazards in facilities, equipment, and processes and is capable of assessing systems from multiple perspectives:

Design

- assessing system design capability to meet user specifications and safety standards
- identifying weaknesses in systems

Physical and operational environments

- Assessing environment to ensure system is appropriately situated, supported, serviced, contained, etc.

Operational and procedural controls

- Assessing engineered controls (ex: automation), sequences of operations, Procedural controls (ex: human interactions) etc.
- Assessing different operational modes – start-up, standby, normal operation, steady & unsteady states, normal shutdown, emergency shutdown, etc.

9.2 Advantages

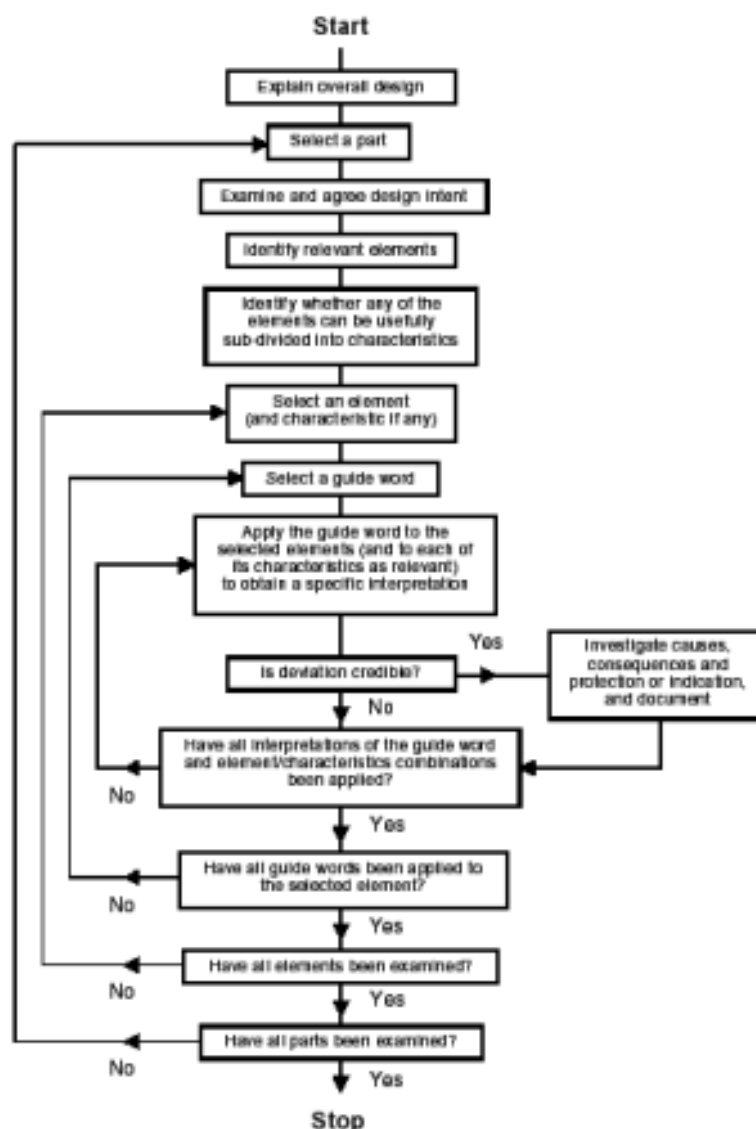
- Helpful when confronting hazards that are difficult to quantify, i.e.:
 - hazards rooted in human performance and behaviours
 - Hazards that are difficult to detect, analyse, isolate, count, predict, etc.
 - methodology doesn't force you to explicitly rate or measure deviation probability of occurrence, severity of impact, or ability to detect
- Built-in brainstorming methodology
- Systematic & comprehensive methodology
- More simple and intuitive than other commonly used risk management tools

9.3 Disadvantages:

- No means to assess hazards involving interactions between different parts of a system or process
- No risk ranking or prioritization capability -teams may optionally build-in such capability as required
- No means to assess effectiveness of existing or proposed controls (safeguards)- may need to interface HAZOP with other risk management tools

9.4 Basic method of HAZOP:

The method applies to processes (existing or planned) for each design information is available. This commonly includes a piping and instrumentation diagram , which is examined in small section , such as individual item of equipment or pipes between them. For each of these a design intention is specified.



9.5 Parameters and guide words:

The key feature is to select appropriate parameter which is applied to design intention. These are general words such as flow, temperature, pressure, composition in order to identify deviation. The current standard words are as follows.

guide words

Meaning

no or not	part of the intent is achieved and nothing else occurs (e.g., no flow)
more	quantitative increase (e.g., higher temperature)
less	quantitative decrease (e.g., lower pressure)
as well as	qualitative increase (e.g., an impurity)
part of	qualitative decrease (e.g., only one of two components in mixture)
reverse	opposite (e.g., backflow)
other than	no part of the intent is achieved and something completely different occurs (e.g., flow of wrong material)

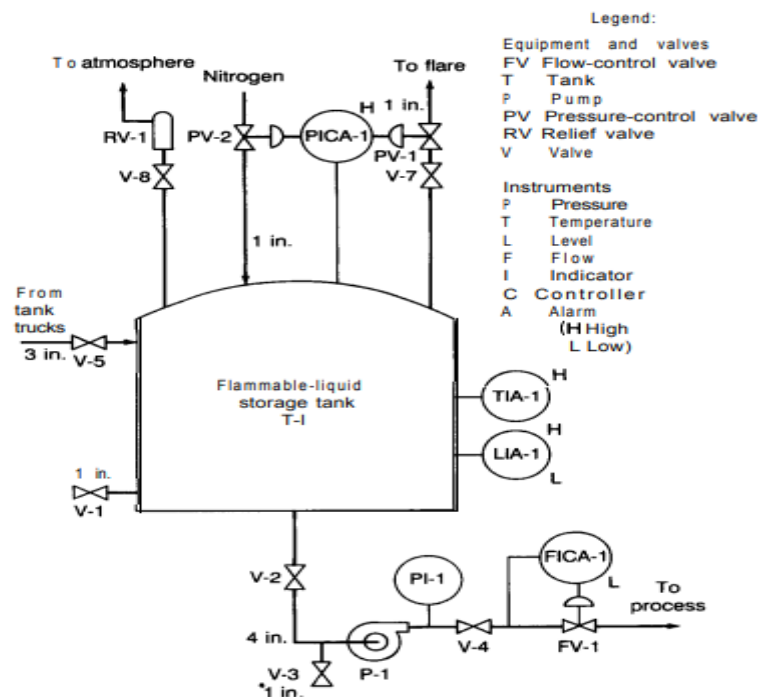


FIGURE 3-1
Piping and instrumentation diagram used in HAZOP example.

9.6 HZAOP study for reactor:

Equipment:Reactor

Parameter: flow

Guide word	Deviation	Cause	Consequence	Action
No	No inlet flow	pump tripping	low quality product	install a micrometer in reactor section
		pump blockage	pressure drop, leakage	regular inspection of transferring lines
More	more inlet flow	feed valve failure and open	increasing unused water	install flow meter at reactor inlet
		leakage in heat exchanger tubes	low quality product	install ratio sensor after mixer
Less	less inlet flow	feed valve failure close	low quality product	regular inspection of transferring lines
		plugging of pipelines	pump damage	install a controller for valve

Parameter: pressure

Guide word	Deviation	Cause	Consequences	Action
high	high pressure	faulty pressure indication	rupture of reactor	high pressure alarm
		vent line closed	effect catalyst activity	continuous check of PSV
low	low pressure	faulty pressure indication	structure disordering of reactor	low pressure alarm

operator error	incomplete reaction	continuous check of PSV
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Parameter: temperature

Guide word	Deviaton	Cause	Consequences	Action
high	high temperature	operator error	chances of reactor melting	high temperature alarm
		failure of temperature indicator	effect reaction rate	emergency shutdown switch
low	low temperature	operator indicator	effect reaction rate	process automation
		failure of temperature indicator	effect on exit gaseous mixture	accurate temperature indicator

10 Plant location

Plant location:

The geographical location of a final plant will have a powerful influence on the success of associate in nursing industrial venture. Tidy care should be exercised in choosing the plant location, and plenty of various factors should be thought-about the selection of ultimate plant ought to be primarily based 1st on a whole survey of the benefits and drawbacks of assorted geographical areas and ultimately on the benefits and drawbacks of accessible realty. The subsequent factors ought to be thought-about in choosing a plant site:

- Raw material accessibility
- Markets
- Energy accessibility
- Climate
- Transportation facilities
- Water offer
- Waste disposal
- Labor offer
- Taxation and legal restrictions
- Site characteristics
- Flood and fireplace protection
- Community factors

Following factors should be thought-about for preparation of PolyEthylene Terephalate to point the requirement for a colossal quantity of data, each qualitative and quantitative.

1. Raw material availability:

It should be assured that staple is well accessible close to the plant as a result of location close to the raw-materials supply permits tidy reduction in transportation and storage charges.

2. Markets:

The location of markets or intermediate distribution centers affects the value of product distribution and therefore the time needed for shipping. It ought to be noted that markets are required for by-products furthermore as for major final product.

3. Energy availability:

Power and steam needs are high in most industrial plants, and fuel is normally needed to provide these utilities. If the plant needs giant quantities of coal or oil, location close to a

supply of fuel offer is also essential for economic operation. The native price of power will facilitate verify whether or not power ought to be purchased or self-generated.

4. Climate:

The climate should be moderate. Severe conditions of temperature rise or decline don't exist during this industrial space. If the plant is found in a very cold climate, prices is also augmented by the requirement for construction of protecting shelters round the method instrumentation, and special cooling towers or air-conditioning instrumentation is also needed if the prevailing temperatures are high. Excessive wetness or extremes of hot or weather will have a heavy impact on the economic operation of a plant, and these factors ought to be examined once choosing a plant website.

5. Transportation facilities:

Transport is a vital issue to think about a plant location that transport isn't therefore costly and time overwhelming.

6. Water supply:

The process industries use giant quantities of water for cooling, washing, steam generation, and as a raw material. The plant, therefore, should be placed wherever a dependable offer of water is out there.

7. Waste disposal:

Proper waste product treatment plant should be gift on the subject of the plant sites. It purifies the waste water from coolers and condensers and releases it to the near waters.

8. Labor supply:

The educated and energized labors are needed for correct operation of plant. The kind and availability of labor accessible within the neighborhood of a projected plant site should be examined.

9. Taxation and legal restrictions:

State and native rate on property financial gain, state insurance, and similar things vary from one location to a different. Similarly, native laws on division, code and transportation facilities will have a significant impact on the ultimate alternative of a plant site.

10. Site characteristics:

Sufficient appropriate land should be accessible for the projected plant and for future growth. The land ought to ideally be flat, well drained and have appropriate supporting characteristics. A full site analysis ought to be created to see the requirement for pillar or alternative special foundations.

11. Flood and fireplace protection:

No flood prone space. It ought to be on the subject of the commercial belt and therefore the plant are made taking preventive measures to fireside hazards and fail safe ways in any contingency.

12. Community factor:

On a brand new site, the area people should be ready to offer adequate facilities for the plant personnel: colleges, banks, housing, and recreational and cultural facilities.

Plant design and economics for chemical engineers 5th edition
by M.S.Peters, K.D.Timmerhaus , R.E.West ch # 02
Chemical engineering design by Richard Coulson
6th edition, 4th volume, ch # 14

11 COST ESTIMATION

Cost engineering is a part of engineering practice related with the “Application of scientific principles and techniques” to solve the problems of:

- Cost estimating
- Cost control
- Business planning and management
- Profitability analysis
- Project management and programming

Cost Indexes

- Prices may change considerably with time due to changes in economic conditions.
- A cost index is merely an index value for given point in time showing the cost at that time relative to certain base time.

$$\text{Present cost} = \text{Original Cost} \times \frac{\text{Index value at present time}}{\text{Index value at time original cost was obtained}}$$

Pumps	Capacity = S	$C_e = a + bS^n$	Cost in 2019
	L/s	\$	\$
Pump 1	6.26	7495	9802
Pump 2	2.87	7685	9232
Pump 3	2.10	7910.5	9328.15
Pump 4	1.89	7580.22	8392.84
Pump 5	0.989	7139.03	8219.9
Pump 6	0.672	8129.69	9702.7
Total Cost	54677\$		

Equipment	Cost(\$)
MIXING VESSEL	76960
DISTILLATION COLUMN	125640
HEAT EXCHANGER	39060
REACTOR	127307
FURNACE	260,982
Pumps	70916
Total	700865

Direct Cost Parameters	% of Equipment Cost	Cost (\$)
Equipment Installation	47 %	329406.6
Instrumentation and Control	36 %	252311.4
Installed Piping Cost	68 %	476588.2

Electrical Component Cost	11 %	77095.15
Building	18 %	126155.7
Yard Improvement	10 %	70086.5
Service Facilities	70 %	490605.5
Total	-	1822249

Indirect Cost Parameters	% of Equipment Cost (E)	Cost (\$)
Engineering and Supervision	33 %	231285.5
Construction Expenses	41 %	287354.7
Legal Expenses	04 %	28034.6
Construction Fee	22 %	154190.3
Contingency Cost	44 %	308380.6

Total	-	1009246
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▪ **Fixed Capital Investment (FCI)**

FCI = Direct Cost + Indirect Cost

FCI = 1822249 + 1009246

FCI = 2.8314 million \$

▪ **Working Capital Investment (WCI)**

WCI = 15% FCI

WCI = 424724.86 \$

WCI = 0.424 million \$

▪ **Total Capital Investment (TCI)**

Total Capital Investment = WCI + FCI

= 0.424 + 2.8314 million \$

Total Capital Investment = 3.2554 million \$

Raw Material Cost

1. **PTA Cost**

Flow rate = 4001 Kg/h

Flow rate = 35049 Ton/yr

Cost = 640 \$/ton = 0.64 \$/Kg

Total Cost = 22.431 million \$/yr

2. **MEG Cost**

Flow rate used = 3.07 Ton/h

Recovered = 1.4 ton/hr

Used MEG = 3.07 – 1.4 = 1.6 ton/hr = 14016 ton/yr

Cost = 856 \$/ton

Total Cost = 11.9976 million \$/yr

- **Operating Labor**

Minimum Wage = 0.4 \$/h

Capacity = 100 ton/day

= 100,000 kg/day

Operating Labor = 50 h/day

Processing Step = 5

Operating Labor = 50 x 5 = 250 h/day

Operating Cost of Labor = 250 x 365 x 0.4

= 36500 \$/yr

- **Total production Cost**

1. **Variable Cost**

Variable Cost = Raw Materials + Miscellaneous Materials

Variable Cost = 34.42 + 0.2 million \$/yr

Variable Cost = 34.62 million \$

Function	Percentage	Cost (\$)
Maintenance	5-10 % FCI	283000
Operating Labor	-	36500
Laboratory Cost	20-23% Operating labor	8030
Supervision	20% Operating labor	8000
Plant Overheads	50% Operating labor	18250
Capital Charges	10% FCI	283000
Insurance	1% FCI	28300
Local Taxes	2% FCI	56600

Royalties	1% FCI	28300
Total	-	749980

▪ **Direct Production Cost**

Direct Production Cost = Variable Cost + Fixed Cost

$$= 34.62 + 0.749$$

$$= 35.36 \text{ million \$ / yr}$$

▪ **Overhead Cost**

Overhead Cost = 30% of Direct Production Cost

$$= 10.61 \text{ million \$ / yr}$$

Manufacturing Cost = Overhead Cost + Direct Production Cost

$$= 10.61 + 35.36 \text{ million \$ / yr}$$

$$= 45.97 \text{ million \$ / yr}$$

▪ **Depreciation**

$$\text{Depreciation} = D = \frac{V - V_s}{N}$$

$$\text{FCI} = V = 2.83 \text{ million \$}$$

$$\text{Salvage Value} = V_s = 5\% \text{ of } V$$

$$\text{Number of Years} = N = 10 \text{ yr}$$

$$D = 0.1415 \text{ million \$}$$

Function	Percentage of Total Production Cost	Cost (million \$)
Administration	2%	0.91

Distribution and Marketing	2%	0.91
Research and Development	5%	2.29
Total		4.11

General Expenses = 4.11 million \$/yr

Total Production Cost = Manufacturing Cost + General Expenses

= 56.988 million \$/yr

▪ **Total Selling Cost Cost**

Flowrate = 100 ton/day = 36500 ton/yr

Cost = 1700 \$ /ton

Total cost = 62.05 Million \$/yr

Gross Income = Total Income – Total Production Cost – Depreciation

= 62.05 – 56.988– 0.19 million \$/yr

Gross Income = 4.87 million \$/yr

Taxes = 40% of Gross Income

Taxes = 2.1 million \$/yr

Net Income = Gross Income – Taxes

= 4.87 – 2.1 million \$/yr

Net Income = 2.71 million \$/yr

▪ **Rate of Return**

$$\text{ROR} = \frac{\text{Net Income}}{\text{Total Capital Investment}} \times 100$$

$$\text{ROR} = \frac{2.71}{3.2554} \times 100$$

ROR= 83 %

- **Payback Period**

$$\text{Payback Period} = \frac{1}{\text{ROR}} \times 100$$

Payback Period = 1.20 yr

13 BIBLIOGRAPHY:

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